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Technology Assessment Report for Industrial Boiler Applications: Fluidized-bed Combustion

GCA Corp.
Bedford, MA

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16. ABSTRACT The report gives results of an assessment of the applicability of atmospheric fluidized-bed combustion (AFBC) to industrial boilers. It is one of a series of reports to aid in determining the technological basis for a New Source Performance Standard for air pollutant emissions from the boilers. It reviews the development status and performance of SO2, NOx, and particulate control options for AFBC; selects the most promising systems for control; and estimates the cost, energy, and environmental impacts of the most promising systems. It concludes that the most promising approach for economically achieving the range of SO2 control levels considered (75 - 90%) involves increased residence time (about 0.67 sec) and decreased sorbent particle size (about 500 micrometers surface mean) in the bed. NOx emissions in the range considered (0.5 to 0.7 lb/10 million Btu) should be achieved in AFBC units without any further control technology. Fabric filters and ESPs appear to be the best options for particle control, although both must yet be demonstrated on AFBC. Cost estimates indicate that AFBC should be able to achieve the levels of control considered, at a cost only moderately above those of an uncontrolled conventional boiler, and at a cost competitive with conventional boilers using flue gas scrubbers.

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i.a

ABSTRACT

This technology assessment report discusses the use of fluidized-bed combustion (FBC) in industrial boilers <73 MW_t (250×10^6 Btu/hr) thermal capacity. The information is being provided to support the industrial boiler control technology assessment study being conducted by the Environmental Protection Agency. The emphasis of the study is on coal combustion. The principles of FBC operation and emission control are identified along with the best systems to meet optional levels of control for SO_2, NO_x, and particulate emissions. The best systems are evaluated based on status of development, performance, cost impact, energy impact, and environmental impact.

Comparison is made with conventional boiler systems, to provide perspective relative to the advantages and disadvantages of FBC. Although AFBC cost and performance remain to be fully demonstrated in commercial application, available data indicate that AFBC should be a candidate for any new coal-fired industrial boiler installation where SO_2 contol is required.

PREFACE

The 1977 Amendments to the Clean Air Act required that emission standards be developed for fossil-fuel-fired steam generators. Accordingly, the U.S. Environmental Protection Agency (EPA) recently promulgated revisions to the 1971 New Source Performance Standard (NSPS) for electric utility steam generating units. Further, EPA has undertaken a study of industrial boilers with the intent of proposing a NSPS for this category of sources. The study is being directed by EPA's Office of Air Quality Planning and Standards, and technical support is being provided by EPA's Office of Research and Development. As part of this support, the Industrial Environmental Research Laboratory at Research Triangle Park, North Carolina, prepared a series of technology assessment reports to aid in determining the technological basis for the NSPS for industrial boilers. This report is part of that series. The complete report series is listed below:

Title	Report number
The Population and Characteristics of Industrial/Commercial Boilers	EPA-600/7-79-178a
Technology Assessment Report for Industrial Boiler Applications: Oil Cleaning	EPA-600/7-79-178b
Technology Assessment Report for Industrial Boiler Applications: Coal Cleaning and Low Sulfur Coal	EPA-600/7-79-178c
Technology Assessment Report for Industrial Boiler Applications: Synthetic Fuels	EPA-600/7-79-178d
Technology Assessment Report for Industrial Boiler Applications: Fluidized-Bed Combustion	EPA-600/7-79-178e
Technology Assessment Report for Industrial Boiler Applications: NO_x Combustion Modification	EPA-600/7-79-178f
Technology Assessment Report for Industrial Boiler Applications: NO_x Flue Gas Treatment	EPA-600/7-79-178g
Technology Assessment Report for Industrial Boiler Applications: Particulate Collection	EPA-600/7-79-178h
Technology Assessment Report for Industrial Boiler Applications: Flue Gas Desulfurization	EPA-600/7-79-178i

These reports will be integrated along with other information in the document, "Industrial Boilers — Background Information for Proposed Standards," which will be issued by the Office of Air Quality Planning and Standards. Therefore, for regulatory purposes, the assessment in this report — and in the companion series of reports — must be viewed as preliminary, pending the results of the more extensive examination of impacts to be conducted by the Office of Air Quality Planning and Standards under Section 111 of the Clean Air Act.

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1.0 EXECUTIVE SUMMARY

1.1 INTRODUCTION

1.1.1 Purpose of the Report

This Technology Assessment Report on Fluidized-Bed Combustion (FBC) has been prepared under contract to the U.S. Environmental Protection Agency (EPA) – Industrial Environmental Research Laboratory (IERL). The information in this report serves as background data for a comprehensive industrial boiler emission control study being conducted by the EPA – Office of Air Quality Planning and Standards (OAQPS). This report, along with several others on emission control technologies will be used by OAQPS to assess the performance of alternative control techniques for industrial boilers.

1.1.2 Scope of the Study

The FBC technology assessment report is a compilation of information gathered from published and unpublished sources and personal communications with FBC manufacturers and researchers, consulting engineers and pollution control vendors. The state-of-the-art regarding the degree of pollution control achievable by fluidized-bed combustion for SO_2, NO_x and particulate emissions is reported. The study analyzes the economic, energy and environmental penalties associated with achieving these emission reductions.

The emphasis of the analysis is on coal-fired units. Despite the fact that fluidized-bed combustion offers multifuel capabilities, the prime interest in the technology is associated with its capability to burn coal efficiently with reduced environmental impact. In addition, the bulk of available operating and experimental data is for coal-firing.

Standard industrial FBC boilers in the size range of 8.8 MW_t to 58.6 MW_t were considered. Commercial FBC units are currently being offered by several vendors across this entire capacity range; in fact, commercial units as small as 2.0 MW_t are now available. AFBC boilers were compared to uncontrolled conventional boilers of the same capacity. This basis of comparison was used in each Technology Assessment Report so that combinations of different boilers and control techniques could be used at a later stage by OAQPS to develop model boiler systems. Three coal types were also considered. Table 1 is a summary of important boiler parameters assessed in this report.

Although fluidized-bed combustion units are offered commercially by several vendors, FBC is still an emerging technology. Most of the currently available data and operating practices are based on bench and pilot scale units. Actual data from commercially operating units are not yet available; hence, it was necessary in some cases to assume a representative range of variables and consider the variables parametrically.

The ranges used in making these assumptions and extrapolations were conservative and the basic conclusions in this report should not change substantially as better data become available.

Fluidized-bed combustion has been deemed by the U.S. Department of Energy (DOE) as one of eight new energy technologies* whose commercialization will

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*Other technologies included in the DOE program are: low Btu-gasification, enhanced oil recovery, unconventional gas recovery, low head hydroelectric power, passive solar energy, energy conserving oil equipment and high efficiency electric motors.

TABLE 1. SUMMARY OF BOILER DESIGN/OPERATING CONDITIONS

Boiler capacity MW_t (10^6 Btu/hr)	Technology	Boiler type	Excess air (%)	Load factor (%)	Coal type and associated operating conditions																	
					Eastern high sulfur						Eastern low sulfur						Subbituminous					
					% S = 3.5 % Ash = 10.6 HHV = 27,447 kJ/kg (11,800 Btu/lb)						% S = 0.9 % Ash = 6.9 HHV = 32,099 kJ/kg (13,800 Btu/lb)						% S = 0.6 % Ash = 5.4 HHV = 22,330 kJ/kg (9,600 Btu/lb)					
					Coal feed rate		Flue gas rate		Flue gas temperature		Coal feed rate		Flue gas rate		Flue gas temperature		Coal feed rate		Flue gas rate		Flue gas temperature	
					kg/sec	(ton/hr)	m^3/sec	(acfm)	°C	(°F)	kg/sec	(ton/hr)	m^3/sec	(acfm)	°C	(°F)	kg/sec	(ton/hr)	m^3/sec	(acfm)	°C	(°F)
8.8 (30)	AFBC	Package watertube/firetube, overbed feed*	20	60	0.32	(1.27)	4.87	(10,300)	177	(350)	0.27	(1.09)	4.61	(9,800)	177	(350)	0.39	(1.56)	4.72	(10,000)	177	(350)
	Uncontrolled Conventional	Package, watertube, underfeed stoker	50	60	0.32	(1.27)	6.09	(12,900)	204	(400)	0.27	(1.09)	5.76	(12,200)	177	(350)	0.39	(1.56)	5.90	(12,500)	177	(350)
22 (75)	AFBC	Partial field erection of shop fabricated modules, watertube, overbed feed	20	60	0.80	(3.18)	12.20	(25,800)	177	(350)	0.69	(2.72)	11.37	(24,100)	177	(350)	0.99	(3.91)	11.86	(25,100)	177	(350)
	Uncontrolled Conventional	Field erected watertube chain grate stoker	50	60	0.80	(3.18)	15.24	(32,300)	204	(400)	0.69	(2.72)	14.21	(30,100)	177	(350)	0.99	(3.91)	14.82	(31,400)	177	(350)
44 (150)	AFBC	Field erected watertube overbed feed	20	60	1.60	(6.36)	24.47	(51,800)	177	(350)	1.37	(5.43)	22.96	(48,600)	177	(350)	1.37	(5.43)	23.71	(50,200)	177	(350)
	Uncontrolled Conventional	Field erected watertube spreader stoker	50	60	1.60	(6.36)	30.58	(64,800)	204	(400)	1.37	(5.43)	28.69	(60,800)	177	(350)	1.97	(7.81)	29.64	(62,800)	177	(350)
58.6 (200)	AFBC	Field erected watertube overbed feed	20	60	2.13	(8.47)	32.59	(69,000)	177	(350)	1.83	(7.25)	30.76	(65,200)	177	(350)	2.63	(10.42)	31.89	(67,600)	177	(350)
	Uncontrolled Conventional	Field erected watertube pulverized coal	30	60	2.13	(8.47)	35.30	(74,800)	204	(400)	1.83	(7.25)	33.32	(70,600)	177	(350)	2.63	(10.42)	34.55	(73,200)	177	(350)

*Overbed feed design is considered because available experimental data indicate equivalent desulfurization performance between in-bed and over-bed feed arrangements as long as primary recycle is practiced (see Section 3.0 and 7.0). Also, the available FBC cost estimates were based on over-bed feed. If in-bed feed is necessary in commercial application to attain high efficiency SO_2 control, the resultant economics are expected to fall within the high error band of the FBC cost estimates presented in Section 4.0.

be accelerated through DOE programs. The data base is expected to expand considerably as more demonstration units come online in the next 2 years.

1.2 SYSTEMS OF EMISSION REDUCTION

1.2.1 Principles of Control

Fluidized-bed SO_2 control technology is based on the reaction of calcium oxide with the sulfur released from coal combustion. A calcium based sorbent, limestone or dolomite, is fed into the bed along with the coal. SO_2 is formed in the bed; the limestone is calcined forming calcium oxide, and the following reaction takes place.

$$CaO + SO_2 + 1/2O_2 \rightarrow CaSO_4$$

NO_x emissions from FBC units resulting from oxidation of organic nitrogen compounds in the coal and thermal fixation of atmospheric nitrogen tend to be low. The mechanisms causing the reduced emissions are not well understood, but are inherent to the fluidized-bed process based on experimental data and observations.

Industrial FBC boilers will generally use a primary particulate control device (a cyclone or multitube cyclone) to recycle 80 to 90 percent (a level achieved in experimentation to date) of elutriated particulate back to the bed. It is expected that flue gas particles downstream of the primary device can be collected at high efficiency by a final control device. Fabric filters, ESPs and multitube cyclones are most applicable.

1.2.2 Control Techniques Considered

A wide cross-section of control techniques was considered for SO_2, NO_x, and particulates. Each of these techniques was assessed in terms of performance (e.g., efficiency, reliability, and versatility); applicability (i.e., compatibility with the full range of FBC industrial boiler capacity); and, status of development (i.e., when the technique would be considered a proven and available technology). The techniques considered are itemized below:

- SO_2 Control
  - Adjustment of Ca/S molar feed ratio
  - Increased gas phase residence times
  - Reduced sorbent particle size
  - Variability of sorbent reactivity
  - Adjustment of bed temperature
  - Variability in feed mechanisms
  - Variability in excess air levels
  - Pressurized fluidized-bed combustion
  - Synthetic sorbents
  - Regeneration of sorbent
  - Enhancement of SO_2 capture with catalysts
- NO_x Control
  - Inherent fluidized-bed combustion chemistry
  - Reduced excess air
  - Increased gas residence time
  - Decreased bed temperature
  - Staged combustion
  - Pressurized fluidized-bed combustion
  - Staged coal feed points
  - Ammonia/urea injection
  - Flue gas recirculation
  - Injection of recycle char
- Particulate Control
  - Fabric filtration
  - Electrostatic precipitation (hot- and cold-side)
  - Multitube cyclones
  - Wet scrubbers
  - Modified design parameters
  - Sorbent treatment to reduce attrition

1.2.3 Degrees of Control Considered

In the ensuing discussion of emission control technologies, candidate technologies are compared using three emission control levels labelled "moderate, intermediate, and stringent." These control levels were chosen only to encompass all candidate technologies and form bases for comparison of technologies for control of specific pollutants considering performance, costs, energy, and nonair environmental effects.

From these comparisons, candidate "best" technologies for control of individual pollutants are recommended for consideration in subsequent industrial boiler studies. These "best technology" recommendations do not consider combinations of technologies to remove more than one pollutant and have not undergone the detailed environmental, cost, and energy impact assessments necessary for regulatory action. Therefore, the levels of "moderate, intermediate, and stringent" and the recommendation of "best technology" for individual pollutants are not to be construed as indicative of the regulations that will be developed for industrial boilers. EPA will perform rigorous examination of several comprehensive regulatory options before any decisions are made regarding the standards for emissions from industrial boilers.

The degrees of control which were considered for current fluidized-bed combustion technology in this assessment are summarized in Table 2.

1.2.4 Best Control Techniques

1.2.4.1 SO_2 Control--

The best system of SO_2 emission reduction is the one which minimizes sorbent feed rates, and still attains high levels of control. The Ca/S molar feed ratio can be reduced with careful control of other operating conditions — most significantly, sorbent particle size and gas phase residence time. Experimental results and theoretical considerations indicate that small particle sizes (in the range of 500 μm) and sufficiently long gas phase residence times (0.67 sec) are representative conditions for effective SO_2 control, although most FBC facilities currently are designed or operated with shorter residence times and coarser sorbent particles. The conditions used in this report for the best system of SO_2 control are:

- Gas phase residence time = 0.67 sec
- Surface average limestone particle size in bed = 500 μm
- Bed temperature = 843°C (1550°F)
- Excess air rate = 20 percent
- Primary recycle of bed carryover

TABLE 2. OPTIONAL LEVELS OF CONTROL TO BE SUPPORTED — ATMOSPHERIC FLUIDIZED-BED COMBUSTION OF COAL

Level of control	SO_2	NO_x	Particulate
	% reduction	ng/J ($lb/10^6$ Btu)	ng/J ($lb/10^6$ Btu)
Stringent	90*	215 (0.5)	12.9 (0.03)
Intermediate	85*	258 (0.6)	43 (0.1)
Moderate	75*	301 (0.7)	107.5 (0.25)

*In addition to the % reduction, an upper limit of 516 ng/J (1.2 $lb/10^6$ Btu) applies in all cases. Furthermore, in no case are controls required to reduce emissions below 86 ng/J (0.2 $lb/10^6$ Btu).

Increased gas residence times and reduced sorbent particle sizes will necessitate reduced gas velocities through the bed, thus increasing boiler size somewhat. It is estimated that especially at elevated SO_2 removal requirements, the possible capital cost penalty associated with the larger boiler will be more than offset by the reduced sorbent and spent solids disposal costs at the recommended conditions.

An important goal in the development of fluidized-bed combustion boilers has been to maximize capacity in a combustion chamber smaller than traditionally possible to allow package fabrication of larger capacity boilers and achieve savings in capital cost. Recommendations in this report concerning "best system" conditions address SO_2 control capability by minimizing sorbent requirements and, thus, enhancing boiler and plant efficiency and minimizing costs associated with sorbent use. The conditions specified above are not much different than those being used in current and envisioned FBC designs. For instance, the design of Combustion Engineering's demonstration boiler (22,700 kg/hr-steam) at the Great Lakes Naval Training Center specifies a nominal superficial velocity of 2.1 m/sec (7 ft/sec), expanded bed height of 0.9 m (3 ft), and in-bed mass mean particle size of 800 μm (which is probably close to a surface average particle size of 500 μm). Considering that early FBC designs called for superficial velocities of 3 to 4.3 m/sec (10 to 14 ft/sec) and in some cases expanded bed depths of less than 0.9 m (3 ft), the conditions recommended do not seem to represent a significant change from currently envisioned nominal design/operating conditions. All of the conditions specified have been used in various experimental programs. FBC technology is still in the development stage so the recommended conditions should be adaptable in future designs.

Some commercially-offered AFBC designs (including over-bed coal feeding and inherent shallow-bed operation) may not be readily adaptable to the increased gas residence time/500 μm particle size conditions recommended here for the best SO_2 control system. Further data on these designs are required to establish their SO_2 control performance.

The selection of increased gas residence time (0.67 sec) and reduced sorbent particle size (500 μm) was made with the use of a mathematical model which can be utilized to project Ca/S requirements based upon laboratory thermogravimetric analysis data. Actual AFBC operating data at conditions near these conditions are limited, but some are available from smaller pilot- and bench-scale units. Therefore, additional data, especially from large AFBC units operating at conditions near the best system conditions are required in order to confirm AFBC SO_2 removal performance at these conditions.

The results of experimentation conducted to data at close to the selected best system conditions were reviewed to assess the correlation between SO_2 removal efficiency and Ca/S ratio. A range of sorbent feed requirements was noted because of differences in the reactivity and capacity of sorbents investigated. The observed ranges in Ca/S ratios are shown in Table 3 for SO_2 removal efficiencies ranging between 75 to 90 percent.

TABLE 3. RANGE OF EXPERIMENTAL Ca/S RATIOS NECESSARY TO MEET OPTIONAL SO_2 CONTROL LEVELS AS OBSERVED IN TESTING AT OR NEAR "BEST SYSTEM" CONDITIONS

Control level	% reduction	Range of Ca/S ratio	Average Ca/S ratio
Stringent	90	2.3 - 4.2	3.3
Intermediate	85	2.1 - 3.8	2.9
Moderate	75	1.6 - 3.2	2.2

The range shown reflects the fact that the impact of the variance in sorbent reactivity on total sorbent needs may override the impact of the optional control levels considered. Other operating conditions (e.g., sorbent particle size, and gas phase residence time) varied slightly in the experimentation used as a basis, but results were screened to maintain such variation to a minimum. Therefore, the sorbent requirements noted in Table 3 represent best SO_2 control, with variation due to sorbent reactivity. This variation is highly probable in the industrial sector because high quality sorbents may not always be available to an individual industry.

Ca/S ratios used by experimenters to achieve 75 to 90 percent SO_2 reduction have been noted as high as 5 or 6. These high sorbent requirements are due primarily to operating factors which were not near best system conditions in combination with a low reactivity sorbent. ANL (the 6 in. diameter unit) B&W (the 3 ft × 3 ft unit) and B&W, Ltd. (the Renfrew unit) all ran tests in which a Ca/S ratio greater than 5 was used. Gas residence times as low as 0.2 sec were used during these tests. Some SO_2 emission data, which are reported for experimentation not conducted at best system conditions are also within the range shown in Table 3. A combination of higher sorbent reactivity and less than optimal operating conditions may produce adequate results. However, performance can be further improved by taking advantage of best system conditions, although slight modifications to current designs would be required.

1.2.4.2 NO_x Control--

The best system of NO_x control capitalizes on the inherent combustion chemistry of the fluidized-bed system. Low temperatures and chemical kinetics combine to produce NO_x emissions which typically are lower than most conventional systems. The levels of control that were considered are shown in Table 4.

TABLE 4. OPTIONAL NO_x CONTROL LEVELS

Control level	Emission rate	
	ng/J	($lb/10^6$ Btu)
Stringent	215	0.5
Intermediate	258	0.6
Moderate	301	0.7

Almost all of the data from experimental AFBC units operating at primary cell bed temperature (<900°C), including units as small as 6 in. diameter, are below the moderate level of 301 ng/J. Essentially all of the data from large AFBC (>500 lb coal/hr), and most of the data from smaller units, are below the intermediate level. The limited data available from the largest AFBC units are consistently below the stringent level of 215 ng/J, although about one-half of the data from smaller units are above that level. Accordingly, it is felt that the stringent level of NO_x control can be achieved in commercial-scale industrial AFBCs, at the values of design/operating variables typically used by process developers today. If the gas residence time is increased for SO_2 control purposes, this may aid in reducing NO_x emissions.

The variables which control NO_x emissions from FBC are not completely understood; thus, it is not possible to define "best" NO_x control options with the same degree of detail that is possible in the case of SO_2. A detailed review of experimental data from AFBC has shown that unit size, bed temperature, excess air, gas residence time, and possibly fuel nitrogen content can influence NO_x emissions, although not with strong, well-defined correlation. The data are sufficiently scattered that it is possible that some minor adjustments to AFBC design/operating parameters may be necessary to ensure that commercial AFBC boilers would achieve the stringent NO_x control level reliably on a 24 hr average basis. Additional data from large AFBC units are necessary to confirm the ability of AFBC to reliably achieve the stringent level without such adjustments. More substantial NO_x control measures (e.g., combustion modifications, such as two-stage combustion) are not felt to be necessary for AFBC to achieve the stringent level of control. Testing of combustion modifications in FBC for improved NOx control is just beginning in some experimental programs.

1.2.4.3 Particulate Control--

The levels of particulate control considered for a fluidized-bed combustion system are shown in Table 5.

TABLE 5. OPTIONAL PARTICULATE CONTROL LEVELS

Control level	Emission rate	
	ng/J	($lb/10^6$ Btu)
Stringent	12.9	0.03
Intermediate	43	0.10
Moderate	108	0.25

Particulate reduction under all three control options should be possible in FBC systems by using conventional add-on particulate control devices. Particle control, adequate to meet these emission levels, has not yet been demonstrated on FBC units, since units of sufficient size have not been operated for sufficiently long periods; however, barring some unexpected unique property of FBC fly ash, it is anticipated that effective control could be achieved by suitable design of conventional particle control devices. The most important factors in selecting a device are reliability and cost. (Other factors are similar for all devices, except environmental impact, where water pollution problems may arise in using wet scrubbers for moderate or intermediate control. Since one of the implicit purposes of FBC is to avoid liquid waste production, use of wet scrubbers is not recommended.)

The control efficiencies required to meet these levels are shown in Table 6.

TABLE 6. CONTROL EFFICIENCIES REQUIRED TO MEET OPTIONAL PARTICULATE CONTROL LEVELS

Fuel and boiler capacity MWt (10^6 Btu/hr)	Particulate emission following primary cyclone ng/J (lb/10^6 Btu)	Particle size average MMD μm	Level of emission control and efficiency of final particulate control device required to achieve that level ng/J (lb/10^6 Btu)		
			Stringent 12.9 (0.03)	Intermediate 43 (0.10)	Moderate 107.5 (0.25)
Coal					
8.8 - 58.6 (30 - 200)	215 - 2150 (0.5 - 5.0)	5 - 20	94 - 99.4	80 - 98	50 - 95

The loadings and particulate size characteristics following the primary cyclone are based on a compilation of experimental results.

Based primarily on the results of the cost analysis, the best devices for stringent and intermediate particulate control should be fabric filters or electrostatic precipitators (ESPs). The best device for moderate control (at collection efficiencies ≤80 percent) should be a multitube cyclone.

The reliability of these systems must be documented in full-scale testing. Experimental data indicate that ESPs will have to be operated as hot-side installations to effectively collect the high resistivity particles elutriated from FBC units. In addition, ESPs may be unreliable for smaller facilities because of possible variations in fuel and sorbent characteristics and the anticipated dependence of ESP performance on these variations.

Fabric filters could possibly have operating problems. Lime hydration at the fabric surface could cause bag blinding. Excessive carbon carryover or temperature excursions could lead to bag fires, even though combustion efficiency in AFBC should be equivalent to well designed conventional stokers.

In any event, potential problems with ESPs and fabric filters must be explored in future commercial scale testing.

1.3 COST IMPACT OF BEST CONTROL TECHNIQUES

Cost estimates for atmospheric fluidized-bed combustion (AFBC) with SO_2, NO_x, and particulate control were developed based on cost quotations from FBC vendors. Costing procedures used by PEDCo for uncontrolled conventional boiler systems[1] were adopted to maintain comparability with those estimates prepared by other TAR contractors for other industrial boiler control technologies. Capital, operating, and total annualized cost were estimated for "grass roots" facilities and the variations based on different levels of emission control were determined. Industrial AFBC boiler cost estimates were also prepared independently by Westinghouse Research and Development and their results are reported for comparison.[2]

1.3.1 Comparison with Uncontrolled Conventional Systems

The cost of AFBC with control was compared with uncontrolled conventional boilers to indicate the cost of control associated with FBC. The accuracy of the results (estimated to be ±30 percent) and validity of conclusions is dependent upon the vendor quotes used as a basis. In certain instances, previous FBC cost estimates were reviewed and reported to lend perspective to the vendor-based estimates.

1.3.2 Cost of SO_2 Control

Costs in terms of $\$/10^6$ Btu output, for industrial AFBC boilers with SO_2 control (excluding final particulate control) are summarized in Table 7. The AFBC costs were developed based on vendor quotations and by employing estimating guidelines recommended by PEDCo early in the program. Uncontrolled conventional boiler costs are shown for comparison and are based on the results of PEDCo's cost analysis.[3] AFBC costs are shown for moderate and stringent SO_2 control with average reactivity sorbent. The worst case FBC cost is also shown; i.e., stringent SO_2 control with low reactivity sorbent.

Considering high sulfur coal, the differential cost between AFBC and uncontrolled conventional systems widens as boiler capacity increases. For stringent control and average sorbent reactivity, the incremental cost for FBC ranges from 4 up to 24 percent of the uncontrolled conventional boiler cost. The worst case incremental costs (low sorbent reactivity) range from 8 up to 30 percent of the uncontrolled conventional boiler cost. The small boiler (8.8 MW_t) costs are roughly comparable due to the simple package design of the FBC unit.

When low sulfur coals are considered, the gap in cost between AFBC and uncontrolled conventional technology narrows due to the significant reduction in sorbent needs and spent solids disposal cost. The 8.8 MW_t AFBC boiler has a slightly lower cost than the comparable uncontrolled conventional boiler. For subbituminous coal, the cost of the two technologies are roughly equivalent at 44 and 58.6 MW_t, even though the conventional boilers are uncontrolled. For other sizes and both low sulfur coals, AFBC technology is roughly 5 to 10 percent more costly than uncontrolled conventional technology.

TABLE 7. SUMMARY OF AFBC BOILER COST WITH SO_2 CONTROL, $\$/10^6$ Btu OUTPUT*†

Coal type	Boiler type	SO_2 control level and % reduction		Sorbent reactivity	Boiler capacity, MW_t			
					8.8	22	44	58.6
Eastern high sulfur	AFBC	Stringent	90	Average	7.75	6.96	5.91	5.69
				Low	8.04	7.28	6.19	5.97
		Moderate	78.7	Average	7.48	6.72	5.65	5.43
	Uncontrolled Conventional		-	-	7.39	5.76	4.77	4.56
Eastern low sulfur	AFBC	Stringent or Intermediate	83.9	Average	6.87	6.21	5.13	4.93
				Low	6.93	6.27	5.19	4.99
		Moderate	75	Average	6.83	6.17	5.10	4.90
	Uncontrolled Conventional		-	-	7.12	5.62	4.70	4.55
Subbituminous	AFBC	Stringent or Intermediate	83.2	Average	6.73	5.88	4.75	4.51
				Low	6.79	5.93	4.80	4.56
		Moderate	75	Average	6.70	5.84	4.71	4.48
	Uncontrolled Conventional		-	-	7.41	5.54	4.73	4.57

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[*] The costs of FBC units with SO_2 control are compared with the costs of uncontrolled conventional boilers in order to provide the incremental cost of using FBC as an SO_2 control system. As indicated in the Preface, similar Technology Assessment Reports have been prepared providing the incremental cost of other SO_2 control options, such as flue gas desulfurization, coal cleaning, and synthetic fuels. A future study by EPA's Office of Air Quality and Planning and Standards will compare the cost of SO_2 removal using FBC and the other control technologies, based upon the Technology Assessment Reports. An initial comparison of controlled FBC with a conventional boiler employing flue gas desulfurization, is included in Section 4.6.2 of this report.

[†] The conclusion suggested by this table — that controlled FBC may be less expensive than uncontrolled conventional boilers in the cases of low sulfur coal — is not supported by some other estimators (Exxon, Reference 4, page i). However, this conclusion is considered to be warranted within the accuracy of the estimates presented in this report.

The costs reported for the 8.8 MW_t (30×10^6 Btu/hr) AFBC are based on a single basic boiler quote. The manufacturer* is currently selling package boilers in this size range. The boiler design is simple, but operates efficiently based on demonstration plant operation over the last several months. Therefore, the costs presented are considered realistic.

The costs for the three larger AFBC boilers are based on quotes from another FBC vendor. This manufacturer* is in an earlier stage of actual commercialization but has been involved in research and development of FBC technology for several years. They are also a major conventional boiler manufacturer.

The cost relationship shown in this analysis indicates AFBC with SO_2 control is generally a higher cost option than uncontrolled conventional technology when field erection is required or when high sulfur coal is burned. Considering all cost estimates (the PEDCo estimates, the independent estimates by Westinghouse, and previous studies by Exxon[4] and A.G. McKee[5]), the values presented for conventional and AFBC boilers are considered to be accurate within 30 percent. Westinghouse estimates of total annual AFBC boiler cost were about 5 percent higher than GCA's for the 8.8 MW_t unit, and about 10 to 15 percent lower for the larger boilers. The difference is in capital cost (direct operating cost estimates were equivalent) but is within the accuracy limits specified. Considering all of these factors it is concluded that, after AFBC costs and performance have been demonstrated, AFBC should be a candidate for any new coal-fired industrial boiler installation where SO_2 control is required.

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*FBC manufacturers are discussed anonymously to maintain confidentiality.

1.3.3 Cost of Particulate Control

The cost of final particulate control in AFBC was assumed to be equal to the cost of final control in conventional boilers burning low sulfur coal. The costs presented in Table 8 are based on vendor quotations and results reported in the TAR on particulate control.[6]

TABLE 8. SUMMARY OF ANNUAL COSTS FOR FINAL PARTICULATE CONTROL DEVICES FOR AFBC INDUSTRIAL BOILERS

Control device	Control level	Annual cost of device, 10^3 $			
		Boiler capacity, MW_t			
		8.8	22	44	58.6
Hot-side ESP	Stringent or Intermediate	63 - 75	147	208	211 - 228
Fabric filter	All optional levels	51	86	147	181
Multitube cyclone	Moderate	10	NA	26	NA

NA = Not available.

The results indicate that fabric filters are the low cost device for stringent or intermediate particulate control, but the estimates assume that there will be no unanticipated baghouse operating difficulties (e.g., bag blinding, bag fires, etc.) that will unduly influence the costs of fabric filter operation on FBC units. The ESP costs are based on hot-side installation to account for noted high particle resistivity in FBC units.

Multitube cyclones appear to be the low cost device for moderate particulate control. Costs were available for ESP use at average SIP levels,* but were all significantly higher than the multitube cyclone costs at a moderate particulate control level. The fabric filters costs shown in Table 8 would not decrease for moderate control because a constant pressure drop has been assumed, regardless of control level.

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*SIP indicates the average emission control level set in State Implementation Plans throughout the United States. For coal, the level is 258 ng/J (0.6 lb/10^6 Btu), a factor of 2.4 more lenient than the optional moderate level under consideration.

The costs presented need to be confirmed in actual application. It is important to emphasize that final particulate control technology has not been demonstrated on AFBC boilers to date.

1.3.4 Cost of NO_x Control

In the large scale AFBC (i.e., B&W 6 ft × 6 ft unit, and Renfrew) NO_x emission testing performed to date, emission levels have not exceeded the optional stringent level of control of 215 ng/J (0.5 lb/10^6 Btu). Additionally, in all testing of smaller bench- and pilot-scale units at temperatures characteristic of envisioned normal AFBC operating temperatures, NO_x emissions have averaged about 215 ng/J (0.5 lb/10^6 Btu). Therefore, it is likely that no special adjustments of FBC conditions will be necessary to achieve the optional levels of NO_x control considered in this report.

If variation of any of the standard design/operating variables (excess air, bed depth, gas phase residence time) were necessary to guarantee reliable (24 hr average) achievement of the stringent NO_x level, there is insufficient correlation in the data to enable rigorous quantification of the cost and effectiveness of parametric variations.

If any adjustments were necessary for NO_x control, it is probable that costs could decrease as well as increase, if such modifications reduce flue gas heat loss or increase combustion efficiency. In fact, any such modifications would be consistent with changes to attain the "best system" of SO_2 control (i.e., increasing gas residence time to 0.67 sec). Further experimentation is required to resolve this effect. For the purpose of this analysis, the costs presented for AFBC boiler operation and SO_2 control are considered to include the cost of NO_x control. No specific costs for NO_x control have been added.

Likewise, the costs of combustion modification techniques to control NO_x (e.g., two-stage combustion) cannot be included because of inadequate data. However, the need for such techniques in FBC, to achieve the NO_x levels under consideration here, is very unlikely.

1.4 ENERGY IMPACT OF BEST CONTROL TECHNIQUES

1.4.1 Basis of Energy Impact Analysis

Energy impact of AFBC commercial application is analyzed with three objectives in mind. These objectives are: (1) quantify the losses in industrial AFBC and conventional coal-fired steam raising equipment sufficiently to permit quantification of energy impact of pollution control; (2) determine total electrical usage for cost estimating purposes; and (3) determine overall boiler efficiency of AFBC and conventional technology for development of cost in terms of $\$/10^6$ Btu output.

To fulfill these objectives each energy loss component was identified. The loss variability was then quantified where possible and the energy loss matrix developed for each component. These components are:

- Coal handling
- Limestone and spent solids handling
- Forced draft, induced draft and other fans
- Boiler water feed and treatment
- Sorbent calcination, sulfation, and spent solids sensible heat
- Flue gas sensible and latent heat losses
- Unburned carbon
- Radiation, convection, and other unaccounted-for losses

1.4.2 Energy Penalty of Air Pollution Control by AFBC

The summation of all energy losses associated with AFBC compared with the losses from uncontrolled conventional boilers are used as the basis for assessing the energy impact of commercialization of AFBC as a control technology. The difference between energy losses in AFBC and conventional technology is defined as the energy impact of control.

1.4.3 SO_2 Control Energy Impact

Because the total of the losses identified in FBC is less than for uncontrolled conventional technology for a capacity of 44 MW_t and below, the energy impact of SO_2 control by AFBC is negative. The energy savings realized by implementation of AFBC over this size range is as high as 3 percent of thermal input based on estimates by GCA; i.e., AFBC boiler efficiency is greater than conventional by as much as 3 percent. The variation is a result of boiler capacity, coal sulfur content, control level and sorbent reactivity. Coal sulfur content has the largest impact, and SO_2 control level appears to have the smallest effect of the parameters considered. If the average SIP SO_2 control level is considered, then the range of SO_2 control is as significant as coal sulfur content in determining energy impact.

When the 58.6 MW_t unit is considered, the uncontrolled conventional unit has lower energy losses than the AFBC boiler with SO_2 control. This is due to greater combustion efficiency in the conventional pulverized coal unit (99 versus 97 percent) and lower flue gas heat losses in the pulverized coal unit than the conventional stokers (30 percent versus 50 percent excess air, respectively). AFBC boiler efficiency at this capacity is 1 to 3 percent lower than that of the uncontrolled pulverized coal boiler. Again, the range results from variation in coal sulfur content, control level, and sorbent reactivity. The range of SO_2 control has a significant effect if the full range of optional levels from SIP to stringent is considered.

Implementation of best system design/operating conditions for SO_2 removal may enhance combustion efficiency by allowing longer carbon residence time in the bed. Also, use of primary recycle allows for combustion of recirculated char.

1.4.4 NO_x Control Energy Impact

No energy impact has been calculated for NO_x control in AFBC boilers. First of all, it is likely that no special FBC system changes will be required to achieve the levels of control being considered; NO_x control would be inherent in the process, and no separate energy impact exists. Second, if some adjustment of FBC design/operating conditions were necessary to achieve the stringent level of control reliably on a 24 hr basis, there is insufficient correlation in the available data to permit quantification of the effect of parametric variations on NO_x emissions. Variables which are known to affect NO_x emissions, but which are not well correlated, are gas phase residence time, excess air, and bed temperature. Other methods of NO_x reduction proposed are two-stage combustion or chemical injection (such as ammonia). When good correlations linking specific parametric variations with NO_x emissions and the effect of these variations on energy loss are developed, energy impact of NO_x control can be properly evaluated, if, indeed, any such parametric variations are necessary to achieve the desired control levels.

1.4.5 Particulate Control Energy Impact

The control methods proposed for FBC particulate control are already commercialized for conventional technology. For the expected dust loadings in

AFBC flue gases, energy use for control will amount to roughly 1 percent of the energy input to the boiler, based upon previous experience with these conventional particle control devices on conventional boilers, burning low sulfur coal.

1.5 ENVIRONMENTAL IMPACT OF IMPLEMENTING BEST SYSTEMS OF CONTROL

1.5.1 Impact of Control Techniques

The major environmental concern in implementing the best candidates for emission control in fluidized-bed combustion is the impact of SO_2 control on the amount of solid waste generated. The amount of spent residue increases as Ca/S ratio is increased to attain higher SO_2 control levels. The major environmental problems with FBC solid waste are high leachate pH, heat release upon initial exposure to water as a result of hydration of the CaO, and total dissolved solids (TDS) above drinking water standards in the leachate.

For perspective, B&W[7] and TVA[8] have compared the amount of waste generated in FBC and conventional boilers using wet, lime/limestone flue gas desulfurization (FGD). Considering plant sizes of 600 and 200 MW_e, respectively, these investigators showed that dry waste amounts were greater for FBC by 10 to 50 percent, but that on a total mass basis (i.e., including the water content of FGD slurry), FGD waste could range as much as 30 percent greater than FBC waste.

Lime/limestone FGD and FBC waste have some similar characteristics in terms of pH, TDS content, and Ca and SO_4 content. However, the following difference has a significant impact. FGD sludge contains sulfite ion ($SO_3^=$) which will be a source of chemical oxygen demand since it is readily oxidized to $SO_4^=$. Whereas FBC waste is dry, and almost fully oxidized, lime/limestone FGD waste is a thixotropic, partially oxidized slurry. Since it liquefies easily it is difficult to handle. Dewatering techniques such as centrifuges and vacuum filters do not reliably yield the 70 to 75 percent solids needed prior to landfilling.

Several other FGD processes appear to be applicable for conventional boiler installations, including sodium scrubbing, double alkali, and Wellman-Lord. All have associated liquid/solid waste streams. In general, solid sludge wastes include calcium and sodium sulfites or sulfates. Liquid wastes from the Wellman-Lord process have low pH and high chlorides. Sodium scrubbing liquid wastes contain about 5 percent solids and sodium sulfates/sulfites or sodium carbonate.

Considering FBC particulate emissions, attainment of high SO_2 control efficiency using high Ca/S ratios and small limestone particle sizes could increase particulate emissions, but it is doubtful that this increase would be to such a degree that available particulate control systems would be inadequate.

Except for the small amount of sorbent which might appear in the fly ash, the quantity of solids resulting from flue gas particle control should be similar to that from a conventional coal-fired boiler.

Implementing the specified levels of NO_x control should require little, if any, change in operating variables and little, if any, environmental impact is foreseen.

It is considered unlikely that combustion modifications (e.g., low excess air, staged combustion) would be necessary for stringent NO_x control. If it were necessary, there could be possible increases in hydrocarbon, CO and particulate emissions, but definitive data are not yet available. Any problems would not be expected to be different than those encountered with combustion modification in conventional systems.

The major environmental impact associated with implementing moderate, intermediate or stringent particulate control is the incremental waste solids/ash to be disposed of.

1.5.2 Solid Waste Disposal

FBC residue does not currently appear to be "hazardous" under RCRA[9] Section 3001, according to the draft procedures currently proposed under Section 3001. Four criteria have been proposed to date for determining whether a material is "hazardous": toxicity (as determined by a proposed leaching test referred to as the Extraction Procedure); corrosivity; reactivity; and ignitability. Several FBC residues have been tested to date under the Extraction Procedure; none were found to be "hazardous" due to toxicity. Also, it is the current judgment that the residue would not be considered corrosive, reactive or ignitable. Therefore, the current conclusion is that FBC residue would generally not be considered hazardous, under the RCRA procedures as currently proposed. Any FBC residue that is found to be hazardous (e.g., due to the use of a particular coal or sorbent having a high trace metal leaching tendency) would be expected to be considered under the "special high-volume waste" category proposed for electric utility residues. Activities are underway by EPA's Office of Solid Waste to expand the RCRA test procedures; biological testing for toxicity is being considered, and a fifth criteria for determining whether a residue is "hazardous" (radioactivity) is under consideration. In addition, changes in the test procedures are possible. These future efforts under RCRA must be followed in order to further assess the status of FBC residue under the Act (and, consequently, any specific disposal requirements that may be imposed).

Potential problems associated with the residue, which have been identified are: the high pH, high TDS, and high Ca and SO_4 in the leachate, the heat release potential upon initial contact with water, and the total solid volume and handling problems.

Solid waste characterization studies indicate that with a judicious choice and design of disposal site, no insurmountable problem should be found. Engineering review of disposal/utilization options, costs, and trace constituents is continuing. Further testing is also needed to assess the biological effects of the leachate from FBC.

1.6 COMMERCIAL AVAILABILITY OF AFBC

AFBC is an emerging technology and commercial sales have just begun. Manufacturers which offer FBC boilers commercially are shown in Table 9. Commercialization is being accelerated by programs sponsored by federal (U.S. Department of Energy) and state (Ohio) agencies to demonstrate the reliability of the systems. Generally the design limestone particle size normally utilized by the companies is higher than that recommended for best SO_2 control systems in this report, and gas residence times are shorter. Thus, SO_2 capture performance may not be as effective for the current designs as projected for best systems. However, FBC systems are flexible, and as more stringent control standards are adopted, it is felt that these variables can be adjusted to come closer to the recommended particle size and gas residence time, without major impact on the FBC process. Only slight modifications in current design/operating specifications would be required. Although increased gas residence time and reduced particle size (reduced gas velocity) will increase boiler capital cost, it is estimated that the reduced operating cost (resulting from reduced sorbent requirements and spent solids disposal) will more than offset

TABLE 9. VENDORS CURRENTLY OFFERING AFBC BOILERS COMMERCIALLY

Company	Location
Fluidized Combustion Company (joint venture of Foster-Wheeler Energy Corporation and Pope, Evans, and Robbins)	Livingston, New Jersey
Johnston Boiler Company (under license to Combustion Systems Ltd.)	Ferrysburg, Michigan
Mustad & Sons	Oslo, Norway
Riley Stoker (with B&W, Ltd.)	Worcester, Massachusetts
Stone-Platt, Ltd.	Netherton, England
International Boiler Works (currently planning to fabricate FBC boilers incorporating designs developed by Energy Resources Company (ERCO), Wormser Engineering, and FluiDyne)	East Stroudsburg, Pennsylvania

the increased capital costs, resulting in reduced steam cost. There is also a possibility of reduced capital costs in other areas, such as particulate control and recirculation pumps (deeper beds may allow for natural coolant circulation). The savings become more substantial as the required degree of SO_2 control is increased. Studies by Westinghouse also support this contention.[10]

Prediction of the nationwide potential for the use of FBC is shown in Table 10,[11] as estimated by EXXON in 1976. Considering that the general industrial boiler market is currently depressed, these estimates may be high. GCA's own investigation indicates that the FBC vendors have the production capacity to build the number of boilers projected for 1985 and 1990, but the demand for this number of installations is uncertain. Implementation of the Fuels Use Act of 1978[12] may have a positive effect on the installation of coal-fired industrial FBC boilers; the law calls for use of coal in new boiler installations (less than 29.3 MW_t) unless technical or economic constraints are prohibitive.

TABLE 10. PROJECTION OF NATIONAL FBC BOILER USE

Year	Cumulative number of industrial FBC boilers	10^{15} Btu per year	1,000 B/D of oil equivalent
1980	7	0.01	5
1985	200	0.29	136
1990	685	0.99	462
1995	1,170	1.69	793
2000	2,050	2.97	1,400

1.7 REFERENCES

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11. Farmer, M.H., et al. op. cit., p. ii.

12. The Fuels Use Act of 1978. Public Law No. 95-620, 92 Stat. 3290.

2.0 EMISSION CONTROL TECHNIQUES FOR FLUIDIZED-BED COMBUSTION

2.1 INTRODUCTION

The main source of air emissions from fluidized-bed combustion (FBC) is the combustion unit itself, and the optional carbon burnup cell, if used. The most important pollutants identified to date are SO_2, NO_x, particulates, and solid residue.

Fluidized-bed combustion provides in situ retention of fuel sulfur and, consequently, lowers the concentration of SO_2 in the flue gas exhausted from the boiler. A suitable bed material such as limestone or dolomite is used to absorb SO_2 formed during combustion. An appropriate Ca/S molar feed ratio (Ca in sorbent versus S in fuel) is selected to meet specific levels of SO_2 removal. SO_2 reduction of 85 percent and higher has been demonstrated in atmospheric fluidized-bed combustion (AFBC), and investigations are continuing to assess the influence of gas phase residence time and sorbent particle size to optimize removal efficiency at low Ca/S molar feed ratios.

Water tubes are submerged directly in the fluidized bed to enhance heat transfer and maintain operating temperatures at 760° to 870°C (1400° to 1600°F). There is experimental evidence that SO_2 removal is optimal in this temperature range.[1] In addition, at this temperature, bed conditions promote the chemical reduction of NO_x formed by oxidation of fuel nitrogen or atmospheric nitrogen.

Uncontrolled NO_x emissions from AFBC are typically in the range of 129 to 258 ng/J (0.3 to 0.6 $lb/10^6$ Btu) at temperatures characteristic of envisioned typical AFBC operation.[2] Current investigations are considering methods of further reduction such as staged combustion, flue gas recirculation, or ammonia injection.

Particulate emissions consist of fuel ash and sorbent elutriated from the bed. Dust loading to the final particulate control device is expected to be similar in quantity to that generated by a conventional system, and will vary depending on fuel ash content, superficial air velocity, sorbent characteristics, the efficiency of primary and secondary cyclones (used for carbon reinjection and preliminary fly ash removal), and whether or not a carbon burnup cell (CBC) is used.

Particulate control in FBC is not thoroughly demonstrated since an FBC unit of sufficiently large size has not yet been operated for a sufficiently sustained period of time. However, the necessary particle control technology for FBC applications should be similar to conventional control applications at conventional boilers burning low sulfur coal. Final particulate capture for an FBC system can be a hot-side or cold-side application (upstream or downstream of final heat recovery) using control devices such as electrostatic precipitators, fabric filters, scrubbers, or cyclones.

2.1.1 System Description — Coal-Fired Fluidized-Bed Boiler

A schematic diagram of an atmospheric pressure fluidized-bed combustion (FBC) boiler is presented in Figure 1, based on a diagram presented by Farmer, et al.,[3] with some modifications by GCA. The unit is comprised of a bed of sorbent (or inert material) which is suspended or "fluidized" by a stream of air at 0.3 to 4.6 m/sec (1 to 15 ft/sec)[4] depending on the density and particle size of the bed materials. Coal, or some other fuel is injected into this

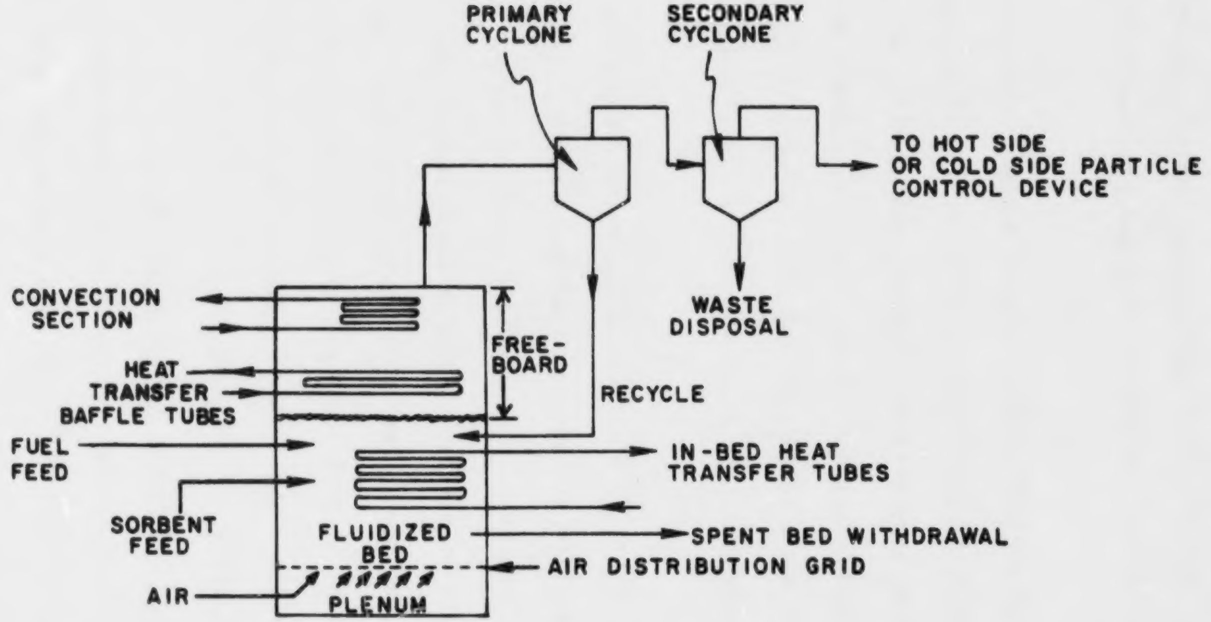
PRIMARY
CYCLONE
SECONDARY
CYCLONE
TO HOT SIDE
OR COLD SIDE PARTICLE
CONTROL DEVICE
WASTE
DISPOSAL
CONVECTION
SECTION
FREE-
BOARD
HEAT
TRANSFER
BAFFLE TUBES
RECYCLE
FUEL
FEED
IN-BED HEAT
TRANSFER TUBES
SORBENT
FEED
FLUIDIZED
BED
SPENT BED WITHDRAWAL
AIR DISTRIBUTION GRID
AIR
PLENUM

Figure 1. Typical industrial FBC boiler.

bed and burned. Sorbent (usually limestone or dolomite) is also injected to react with the SO_2 formed upon combustion. The gas velocity is set so that the bed particles are suspended and move about in random motion. Under these conditions, a gas/solid mixture behaves much like a liquid (e.g., seeks its own level, can be readily moved through channels). The boiler tubes submerged in the bed remove heat at a high rate to maintain bed temperatures in the range of 760° to 870°C (1400° to 1600°F).

Bed material consists of particles with a maximum size of about 0.6 cm (1/4 in.), and is comprised of reacted and unreacted sorbent (limestone, dolomite), ash and other inert material, and small quantities (less than 3 percent) of unburned carbon.[5] The air and combustion gases passing through the bed entrain particles into the freeboard section of the boiler, or carry some of the smaller particles completely out of the boiler. Boiler tubes can be placed within the freeboard for convective heat transfer and also to act as baffles to contain some of the entrained particulate.

Particulate matter completely elutriated from the boiler passes to a primary cyclone where 80 to 90 percent of the larger carbon containing particles are removed.[6] This collected material can be recirculated back to the FBC unit, fed to a carbon burnup cell (CBC) to maximize combustion efficiency, or disposed of. A carbon burnup cell is a separate FBC reactor which is operated at higher temperatures (1093°C (2000°F)) than the main FBC to achieve maximum carbon utilization. A secondary particle collector can be installed to collect fly ash for disposal.

Final heat recovery can be achieved in an economizer and/or air preheater. Final particulate collection (i.e., after primary and/or secondary cyclones) can be achieved either upstream (hot-side) or downstream (cold-side) of final heat recovery.

2.1.2 Mechanisms for SO_2 Control

Sulfur dioxide emissions are a major problem in conventional coal-fired industrial boilers. However, by using FBC technology, SO_2 emissions can be reduced by up to 90 percent or more depending upon the rate of sorbent addition to the bed and the FBC design and operating conditions. The coal is burned in the bed in the presence of lime (CaO). The SO_2 reacts with the calcium oxide and excess oxygen forming calcium sulfate ($CaSO_4$).[7]

$$SO_2 + CaO + 1/2\ O_2 \rightarrow CaSO_4 \text{ (anhydrous)}$$

The CaO in the reaction is produced by rapid calcining of calcium carbonate. The sorbent is most commonly limestone or dolomite. The degree of SO_2 capture possible in FBC industrial boilers is strongly dependent on the calcium to sulfur molar feed ratio (Ca/S). Other factors which affect the sulfur capture efficiency of the system are the reactivity of the sorbent, the particle size of both sorbent and coal, gas residence time in the bed (determined by superficial gas velocity and bed height), the feed mechanism and material distribution in the bed, and temperature. These parameters can be adjusted to obtain the maximum SO_2 removal for the system at a particular Ca/S molar feed ratio.

SO_2 control will be achieved typically on a once-through basis. In a once-through system, spent sorbent is removed from the combustor and disposed of as sulfated stone. Although sorbent regeneration will not likely be used in the near future in industrial FBC boilers, a typical regeneration technique would process the spent stone in a separate reaction vessel by reductively decomposing the spent sorbent to form CaO and SO_2. The SO_2 would be sent to a sulfur recovery system to generate elemental sulfur or sulfuric acid. The regenerated stone as CaO could then be recycled to the combustor as makeup sorbent.

2.1.3 Mechanisms for NO_x Control

Nitrogen oxide (NO_x) emissions from FBC are inherently lower than uncontrolled emissions from conventional combustion. The primary reason for this seems to be the unique combustion chemistry which occurs in the fluidized bed. The fact that the combustor temperature is considerably lower in FBC (815° to 930°C (1500° to 1700°F)) than conventional combustion (1500°C (2700°F)) also aids in lowering NO_x emissions due to reduced fixation of atmospheric nitrogen, but does not seem to be the predominant factor. Formation of NO_x at the lower temperatures is primarily due to the oxidation of fuel nitrogen.[8]

$$2N\ (\text{fuel}) + O_2 \rightarrow 2NO$$

The NO is formed rapidly as the coal burns and is thought to be reduced in the presence of carbon monoxide and other products of incomplete combustion, by a reaction such as the following:[9]

$$2CO + 2NO \rightarrow 2CO_2 + N_2$$

At higher conventional combustion temperatures a larger proportion of NO_x is derived from the oxidation of atmospheric nitrogen:[10]

$$N_2\ (\text{atmospheric}) + O_2 \rightarrow 2NO$$

The reaction rate is relatively slow and temperature dependent. The temperature and the NO_x residence time are not conducive to the NO reduction reaction, so that the final NO_x emissions from conventional boilers are higher than those from FBC.

Some combustor design and operating conditions tend to increase NO_x emissions; e.g., increasing bed temperature, increasing excess air, decreasing gas residence time, and possibly increasing fuel nitrogen content. However, the influence of these variables on NO_x emissions cannot be quantitated or correlated; the mechanisms of NO_x formation and decomposition in FBC are not well understood.

Experimental NO_x emissions data are scattered. Hence, it is not possible to design FBC's for low NO_x emissions with the same reliability possible for SO_2.

Combustion modification methods which are used to reduce NO_x emissions in conventional boilers can also be applied to fluidized-bed combustion. Preliminary experimentation indicates that staged combustion may be successfully applied to FBC.[11] The bed would be operated at low excess air, which inhibits the formation of NO_x. Secondary air would then be injected above the bed to complete the combustion process. Further investigation on large-scale FBC units is necessary to confirm the benefit of implementing combustion modifications on industrial FBC boilers.

2.1.4 Mechanisms for Particulate Control

Particulate matter emitted from the combustion section of an FBC coal-fired boiler consists of fly ash from the coal, unburned carbon, and elutriated sorbent material. (Most of the spent sorbent will be withdrawn from the bed as a solid residue, and, thus will not appear in the flue gas, except in the case of advanced FBC concepts involving high-sorbent-recirculation techniques.) The superficial gas velocity is an important factor in determining particulate escape from the combustor. A high percentage of small-sized particles with terminal settling velocities less than the superficial air velocity will be blown out of the bed. Due to turbulence in the system, geometry, and freeboard height, some larger particles will also be elutriated, and some small particles will remain.[12] The amount of sorbent particulate matter passing out of the bed will depend upon particle size reduction brought about by attrition and decrepitation, which refer to particle grinding and roasting, respectively.

A primary cyclone is used to collect larger particles containing the most significant carbon concentration for circulation back to the FBC or to a separate carbon burnup cell (CBC). A secondary cyclone of higher efficiency can also be used to collect smaller particles for disposal as ash. Design of combustors with high freeboard or baffle heat exchange tubes in the freeboard can help to reduce the amount of particulate elutriated to the primary cyclone.

Final particulate control (after primary and/or secondary cyclones) will be provided by use of conventional systems such as electrostatic precipitators, fabric filters, scrubbers, or cyclones. These systems can be operated as hot-side or cold-side units (upstream or downstream of final heat recovery), except for fabric filters which must be installed cold-side to prevent fabric burning. Although no final stage particulate control device has yet been demonstrated on an FBC unit, it is expected that, by suitable control device design and operation, conventional particle control devices should be adequate to meet the optional emission levels considered in this study.

ESPs are a demonstrated control device on large conventional combustion units, and are capable of removing small particles (<5 μm) at high efficiency. However, resistivity of the particulate from FBC units is expected to be high, due to lime, limestone, and calcium sulfate in the flue gas and low concentrations of SO_2. If current problems with high particle resistivity can be overcome, ESPs may be used on FBC industrial boilers.

Fabric filters have been demonstrated for utility boiler applications, and may be especially applicable for industrial FBC particulate control because of high collection efficiency and insensitivity to particle resistivity. Due to low SO_2 concentrations and a low acid dew point in FBC flue gas, a fabric filter could be operated at low temperatures without fabric deterioration. Potential problems with fabric filter application in FBC include blinding and bag fires. Blinding could occur depending on flue gas moisture and the possibility of calcium oxide hydration in the baghouse. The potential also exists for bag fires if unburned carbon loadings become excessive and temperature excursions occur in the baghouse during transient conditions such as startup or shutdown.

Scrubbers could be used, but pressure drops required for high efficiency small particle removal may be excessive. In addition, the potential for a water pollution control problem exits.

Cyclones may not be capable of providing satisfactory retention of small particles <5 μm. However, they may be used in the smaller boiler size categories depending on control level required because of potential overall system cost advantages. Application of more sophisticated devices on small capacity FBC boilers may result in an unwarranted economic penalty to the industry. The effectiveness of multitube cyclones, cyclones which operate at high differential pressure, or advanced cyclone designs, needs to be explored. In general, further study is required to determine the most appropriate final particulate collection method for FBC systems of different size firing different fuels.

Fly ash handling requirements will be similar to conventional combustion system needs. The major additional equipment needed for FBC system operation is sorbent feed and spent sorbent handling facilities. An advantage of FBC systems is that spent stone can be handled in dry form. Coal feeding may also be different in FBC, especially if underbed feeding is used. This technique would use air injectors to spread the coal throughout the volume of the bed. In bed feeding may be needed to provide suitably long sorbent residence time for highly efficient SO_2 control. To date, experimental results indicate that primary recycle should be capable of providing the necessary residence time, but further work is necessary to confirm this.

Another major equipment need is the forced draft fan which has to overcome approximately three times the pressure drop encountered in a conventional boiler. The additional pressure is needed for air passage through the distribution plate and for bed fluidization.

2.1.5 Differences in Possible AFBC Industrial Boiler Designs

Several alternative AFBC industrial boiler designs are possible. Table 11 summarizes potential alternative generic boiler arrangements. Table 12 lists specific design differences among vendors that are developing FBC boilers for commercial offering. The design conditions which impact emission control are noted. Most industrial AFBC boilers currently offered are designed with water tube heat exchangers in the bed. Additional heat transfer surface in the freeboard can also be used. The Johnston Boiler Company is offering a combined water tube/fire tube unit as shown in Figure 2.[13] The Battelle Multisolids Fluidized Bed uses a separate ancillary dense bed for heat exchange and an entrained bed for combustion.[14] A fluidized-bed air heater is offered by the FluiDyne Company.[15]

2.1.5.1 Coal Feed Systems--

Different coal feed mechanisms are being used by different manufacturers. Stone-Platt is manufacturing systems in which the coal is screw fed just below the top surface of the bed at the center of the unit.[16] The demonstration unit under construction at Georgetown University (designed by Foster-Wheeler and Pope, Evans and Robbins) will utilize an overbed spreader coal feed system.[17] AFBC boilers using staged combustion are offered by O. Mustad and Sons of Gjovik, Norway.[18]

TABLE 11. SUMMARY OF POTENTIAL ALTERNATIVE AFBC INDUSTRIAL BOILER SUBSYSTEM DESIGNS

Subsystem	Possible alternatives	Comments
Fuel and sorbent feed mechanism	In bed or above bed (by vibrational, pneumatic, or stoker feeding)	The air pollution impact of overbed feed AFBC systems is unkown. It is anticipated that SO_2 and NO_x emissions may be increased with overbed feed systems.
	Single point versus multiple point injection	Multiple point injection generally results in better bed mixing.
Heat transfer	Water, steam, air and other media	To date, only water, steam, and air have received much consideration.
	In bed or above bed or both; or in separate ancillary bed	Heat transfer surface in the AFBC freeboard can be water tubes or fire tubes. Battelle Multisolids Unit is using separate ancillary bed for heat exchange.
Bed type	Deep or shallow	Deep bed is usually in the range of 1 meter (3 to 4 feet). Shallow beds of about 0.3 meters (6 to 12 inches) are proposed for use in staged combustion.
	Dense, lean, or entrained	Dense bed operated at low gas velocity provides best emission control. Lean bed operated at high gas velocity to provide good mixing and high heat transfer.
Elutriated solids	Disposed of as ash or recirculated to main bed or carbon burnup cell	Recirculation is being considered to improve combustion efficiency and SO_2 capture.
Spent bed material	Direct disposal or regeneration with recycle to main bed	Regeneration of sulfated stone is being investigated to minimize sorbent makeup and disposal rates.

TABLE 12. DESIGN/OPERATING CONDITIONS OF "COMMERCIALLY-OFFERED" AFBC INDUSTRIAL BOILERS

Design/operating conditions	Foster-Wheeler (Georgetown design)	Babcock & Wilcox (6' × 6' pilot plant)	Combustion Engineering (Great Lakes design)	Johnston Boiler Co. (demonstration boiler)	FluiDyne air heater (3.3' × 5.3' vertical slice combustor)	Stone Platt		B&W, Ltd. (Renfrew design)	O. Mustad (Enköping*)
						Hot water	Steam		
Bed parameter									
Temperature, °F	1,594	1,550	1,550	1,550	1,325 - 1,405	-	-	1,560	1,560
Expanded bed depth, ft	4.5	4.0	3.0	2.67	3.5 - 4.0	-	-	2.6 - 3.0	0.8 (slumped)
Gas velocity, ft/sec	8	8	7	6	2 - 6.0	-	-	8.0	8.2
Gas residence time, sec	0.56	0.50	0.63	0.44	0.6 - 2.0	-	-	(~0.35)	-
Freeboard height, ft	-	-	-	4	-	-	-	-	-
Number of cells	2	5	-	3	-	-	-	-	
Bed area, ft²	11 × 19.3	22 × 26	8 × 17	6 × 4.3	-	4 × 4	8 × 13	10 × 10	10 × 10
Heat trans. coef, Btu/hr ft² °F	40 - 50	-	40	-	-	-	-	70 - 80	-
Combustion efficiency, %	97	-	-	-	-				
Steam/water/air conditions									
Rate, lb/hr	100,000	200,000	50,000	10,000	1,000 - 12,300	30,000	25,000	40,000	-
Pressure, lb/in.²	675	150	365	135	-	-	200	400	-
Temperature, °F	499	358	560	350	-	180	-	560	374
Feedwater, °F	228	240	230	212		140	140	-	284
Feed Conditions									
Coal									Polish
Feed rate, lb/hr	9,565	-	6,710	-	150 - 900	240	3,000	-	6,000 - 10,000‡
HHV, Btu/lb	12,750	11,000	10,430	-	-	10,850	10,850	-	10,800
Sulfur, %	3.3	3.0	3.5	-	3.6	-	-	3.0 - 5.5	1.2
Size	1" × 1/4"	1/4" × 0	1-1/2" × 0	1-1/4" × 0	-	1-1/2" × 0	2-1/2" × 0	-	~1.4"
Feeder type	Overbed	Underbed	Underbed	Overbed	Underbed	-	-	Underbed†	-
Sorbent	Greer/Grove	-	-	-	Owatonna Dolomite	-	-	High-low reactivity	Sala Dolomite
Feed rate, lb/hr	3,133	6,500	3,260	-	50 - 180	-	-	-	-
Ca/S	3	4	3	-	1.1 - 2.2	-	-	2.5 - 5.0	-
Size	4 Mesh	3/8" × 0	-	85% > 16 Mesh, 100% < 8 Mesh	<1/4"	-	-	-	-
S capture, %	90	90	85+	-	-	-	-	90	-
Air									
Excess air	20	21	20	25	30 - 130	-	-	20	10
Boiler efficiency	83	86	82	81	-	~80	~80	-	-
Flue gas temperature, °F	400	300	-	305	-	-	-	-	-
Air pollution control	Baghouse	Same	Same	Baghouse	-	-	-	Prat-Daniel precipitator	Bahco baghouse
Impact of design conditions on emission control	Overbed feeding may not allow for sufficient sorbent residence time to obtain high efficiency SO_2 removal. Early experimental results indicate that primary cyclone recycle can override this potential problem, but this must be confirmed with further investigation.	Reduction of gas velocity to 6 ft/sec would be consistent with best conditions recommended for SO_2 control in Section 3. This would provide for gas residence time of approximately 0.67 seconds.	Gas velocity is high and gas residence time is low as a result. This impacts SO_2 removal performance.	Overbed coal and sorbent feed may not allow for sufficient sorbent residence time to obtain high efficiency SO_2 removal. Primary recycle by knock-out baffle may also provide performance inferior to primary cyclone.	Design conditions are good for SO_2 control. Gas residence time ranges between 0.8 to 2 seconds. Underbed or overbed feed is used. The long residence time may also allow for enhanced NO_x control.	Although dimensions are not available, the vendor claims that the concave shape of the bed bottom causes in bed solids circulation and provides long particle residence times which allows for good desulfurization performance.		The low gas residence time of 0.36 seconds is the main factor which could adversely affect SO_2 control.	This design has a very shallow bed which could impact SO_2 control, but no data is available from this unit to verify this possibility. This design also employs two-stage combustion which could enhance NO_x control.

---

* [illegible] MW output (heat transferred to water).

† [illegible] feed points are used

‡ By screw conveyor (1 point per 6 sq. ft)

Capable of 110 g/m³ (50 gr/ft³)

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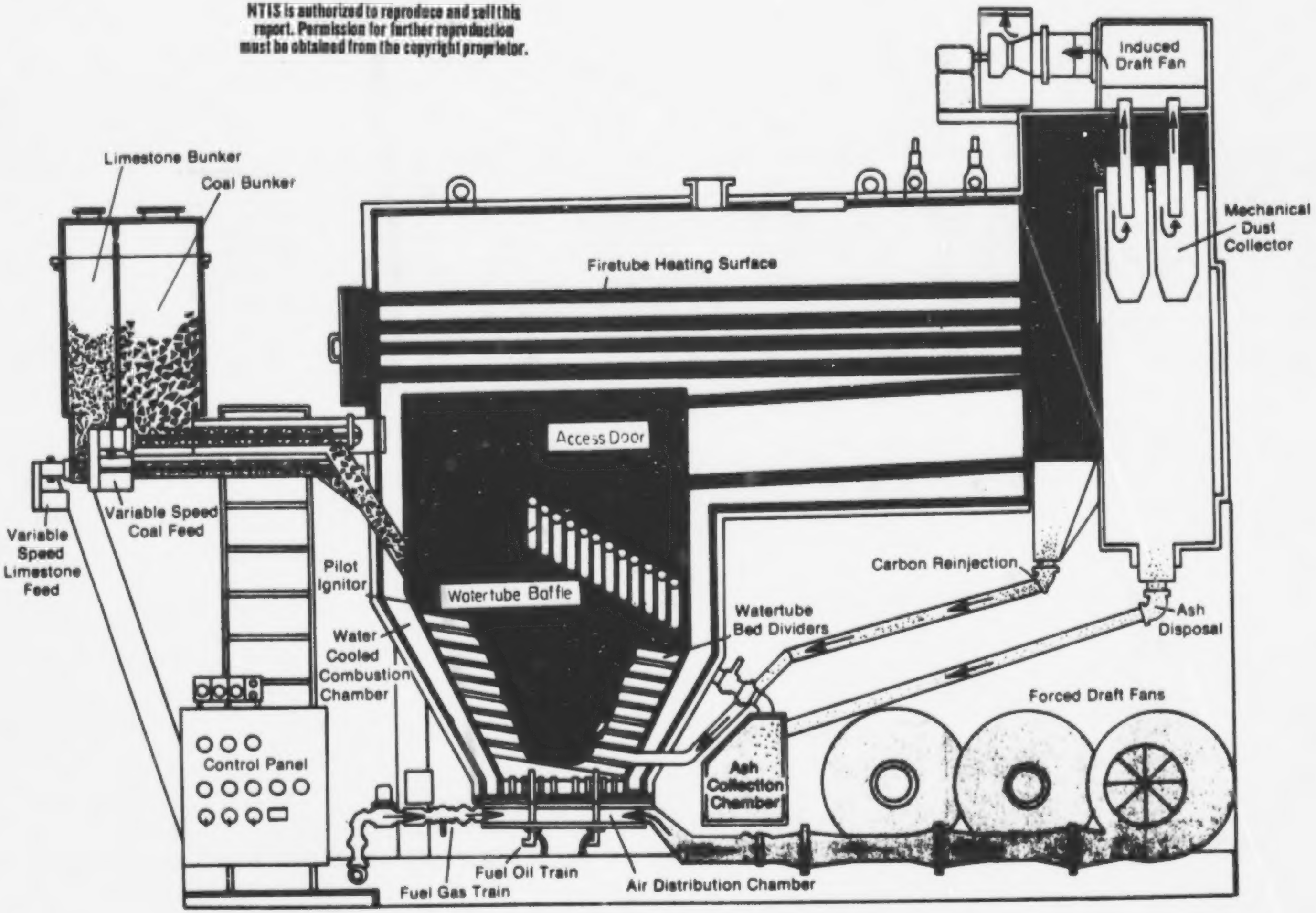

Limestone Bunker
Coal Bunker
Induced Draft Fan
Firetube Heating Surface
Mechanical Dust Collector
Access Door
Variable Speed Coal Feed
Variable Speed Limestone Feed
Pilot Ignitor
Watertube Baffle
Water Cooled Combustion Chamber
Carbon Reinjection
Watertube Bed Dividers
Ash Disposal
Forced Draft Fans
Control Panel
Ash Collection Chamber
Fuel Oil Train
Fuel Gas Train
Air Distribution Chamber

Figure 2. Johnston Boiler Company's combination watertube/firetube FBC boiler.[13]
(Reproduced with permission.)

2.1.5.2 Solids Handling and Disposal--

Most experimental, demonstration, and commercially available systems incorporate recycling of elutriated bed solids to maximize combustion efficiency. Small boilers (less than 15 MW_t) will generally recycle elutriated solids to the main bed while larger systems may recirculate to a separate carbon burnup cell. The Rivesville plant constructed by Foster-Wheeler is a multicell unit which includes a carbon burnup cell.[19] The boiler at Georgetown University is designed with two cells, one of which can be used as a duplicate main cell or for burning recycled material.

In first generation AFBC boilers spent bed material will be withdrawn for direct disposal or byproduct recovery. Regeneration is a long-term development which will find greatest application in utility boilers or in industrial parks as a means of reducing sorbent feed and disposal requirements.

2.1.6 Impact of Key Design Features

Key features which could impact emission control performance in these commercial designs are method of solids feed (overbed feed or underbed feed), bed depth, superficial gas velocity, and sorbent particle size.

2.1.6.1 Superficial Velocity--

Most of the existing designs employ some combination of superficial velocity and bed depth which allows for gas residence times of 0.5 sec or less. The notable exception is the FluiDyne design, which for the conditions listed in Table 12, attains gas residence times between 0.6 and 2 sec. Gas residence times of 0.5 sec and below may require unnecessarily high Ca/S ratios to attain high desulfurization levels. This impacts energy efficiency, overall system cost, and waste disposal. NO_x control may also be slightly limited at lower gas residence times. Estimated best conditions for bed depth, superficial velocity and gas residence time are discussed in Section 3.0.

2.1.6.2 Coal and Sorbent Feed Mechanisms--

Solids feed orientation can also affect emission control. Overbed feeding is technically simpler than underbed feeding, but solid and gas residence time may be less than desirable. SO_2 released above the bed would be captured with reduced efficiency and sorbent may be elutriated before it has a chance to react. There is early indication from FluiDyne testing that feed method may be of minor importance in SO_2 control as long as primary recycle is practiced. However, this needs to be confirmed in more extended large-scale testing.

2.1.6.3 Particle Size--

Another design parameter of major concern with respect to SO_2 control is sorbent particle size. The feed particle size distributions noted by the vendors suggest inbed average sizes in the range of 1,000 to 2,000 µm. This cannot be estimated with certainty because only the top and bottom size limits of the feed sorbent are noted, and the extent of particle attrition in the bed is unknown. However, experimental data and theoretical considerations suggest that inbed particle sizes of about 500 µm surface average are appropriate for good SO_2 control. Overall sorbent requirements can be reduced by using smaller particles with primary recycle.

The Mustad system is worthy of note since it is designed with a shallow bed and two-stage combustion. Although this may provide significant reduction of NO_x, the impact on SO_2 control must be verified.

2.2 STATUS OF DEVELOPMENT

Fluidized-bed combustion is an emerging technology for the clean combustion of fuels. First experimentation with FBC for steam generation was conducted by Combustion Engineering, Inc., in the early 1950s, the British in the early 1960s, and PER in the mid-1960s under sponsorship of the Office of Coal Research.

2.2.1 U.S. Department of Energy Development Programs

The U.S. Department of Energy (DOE) Office of Fossil Energy is conducting an extensive program for development of coal-fired industrial AFBC boilers as part of the National Energy Research, Development, and Demonstration Program, to fulfill the following objectives:

- Identify and conduct evaluations of industrial boiler or process heater requirements to determine the applications in which FBC is technically, economically, and environmentally feasible.
- Obtain sufficient data from prototype operations to design and construct a commercial-size unit.

Four FBC demonstration units are currently in the design or construction phase as a result of ongoing DOE Programs.

The four units are being developed by:

- Combustion Engineering
- Fluidized Combustion Company (joint venture of Pope, Evans, and Robbins, and Foster-Wheeler)
- Battelle Memorial Institute
- EXXON Research and Engineering Company

2.2.1.1 Combustion Engineering - Great Lakes Naval Training Center--[20]

Combustion Engineering will develop a package fabricated coal-fired industrial steam generation boiler. Their work is divided into two phases. The first is design and construction of a subscale test unit with a bed area of 0.3 m^2 (3.0 ft^2) capable of generating 1,044 kg/hr (2,300 lb/hr) steam. This unit is currently operating. The second phase is design and construction of a commercial-scale FBC package boiler capable of generating 22,700 kg/hr (50,000 lb/hr) steam with a coal feed rate of 2,270 kg/hr (5,000 lb/hr). This unit will be located at the Great Lakes Naval Training Center in Illinois and is scheduled for startup in 1981.

2.2.1.2 Foster-Wheeler/Pope, Evans, and Robbins - Georgetown University--[21]

Foster-Wheeler and Pope, Evans, and Robbins are jointly completing installation of a 45,400 kg/hr (100,000 lb/hr) steam generating FBC on the campus of Georgetown University in Washington, D.C., which will supply steam for space heating at the University. Startup began during the summer of 1979.

2.2.1.3 Battelle — Multisolid Fluidized-Bed Combustion (MSFBC)--[22]

The Multisolid Fluidized-Bed Combustion process was developed under the Battelle Energy Program over a 3-year period. The feasibility of this concept has been successfully demonstrated in a 6 in. diameter coal-combustion unit. The U.S. DOE contract with Battelle calls for a two-phase scale-up of this process over 6 years (including 3 years of operating the demonstration plant). The Sub-Scale Experimental Unit System (SSEUS), which represents a 10-fold scaleup of the 6 in. bench-scale unit, is now in operation. This pilot-scale unit is designed to produce about 1,820 kg/hr (4,000 lb/hr) steam from 182 kg/hr (400 lb/hr) coal. The full-scale demonstration plant, which will be built adjacent to Battelle's present steam plant, will represent a further scale-up of about six times and will produce 11,350 kg/hr (25,000 lb/hr) steam while burning 1,135 kg/hr (2,500 lb/hr) coal. Data obtained from this demonstration unit will be used to design and build commercial boilers. The MSFBC consists of a combined dense and entrained fluidized bed to accomplish combustion and desulfurization. Entrained bed material can be recirculated to the dense bed.

2.2.1.4 EXXON - Crude Oil Heating System--[23]

Some proportion of crude oil (~4 to 12 percent) processed in an oil refinery is consumed to maintain refinery operations. Under DOE contract, the EXXON Research and Engineering Company is exploring the feasibility of using coal combustion processes to satisfy this energy requirement. The objectives of the program are first to extend the state-of-the-art of fluidized-bed crude oil heating for refinery applications. Second, an FBC indirect-fired process heater will be designed and constructed as an integral part of a petroleum refinery. Phase I of the program includes the following three laboratory experiments:

- Two dimensional flow visualization units
- Process stream coking unit
- High temperature heat flux unit

Phase II incorporates installation and demonstration of a coal-fired FBC process heater at an EXXON refinery with a capacity between 2.9 to 4.4 MW_t (10 to 15 × 10^6 Btu/hr).

2.2.1.5 Anthracite Culm Combustion Program--[24]

The anthracite culm combustion program was developed by DOE based on successful results at the Morgantown Energy Research Center. Three demonstration units are planned in the State of Pennsylvania as follows:

- City of Wilkes-Barre

  Foster-Wheeler and Pope, Evans, and Robbins will build a 45,400 kg/hr (100,000 lb/hr) FBC boiler burning an anthracite coal/culm mixture to produce steam for district heating and air conditioning within the city. Fuel will be obtained from the Pine Ridge Anthracite bank located in the city. The City of Wilkes-Barre is the prime contractor and program administrator. Foster-Wheeler is responsible for hot model testing and boiler design and erection. Pope, Evans, and Robbins will provide overall system layout, detail design, and program management.

- Shamokin Area Industrial Corporation (SAIC)

  A 9,080 kg/hr (20,000 lb/hr) FBC boiler burning anthracite culm will be installed at the Cellu Products paper reprocessing plant in Shamokin. Fuel will come from the nearby Swift Colliery. SAIC is the prime contractor responsible for site selection, feedstock supply, and steam user coordination. Other contractors involved are Curtiss-Wright, Dorr-Oliver, and Stone and Webster. Curtiss-Wright will provide overall program management. Dorr-Oliver will conduct subscale testing, process selection, and assess prototype performance. Stone and Webster will provide architectural/engineering services, including equipment design and selection, specification and bid package preparation, and assessment of environmental control.

- FluiDyne Engineering Company

  FluiDyne, together with Deltrak and Nebraska Boiler Company will install a boiler at the GTE Sylvania plant in Towanda, as a replacement for an existing oil-fired boiler. The unit will generate 9,080 to 13,600 kg/hr (20,000 to 30,000 lb/hr) steam. FluiDyne is the prime contractor responsible for all subscale testing, engineering, procurement, and construction. The boiler package will be subcontracted through the other two firms mentioned above.

2.2.1.6 Recent Drive for Accelerated Commercialization--

As of April 1979, DOE continued its commercialization drive for industrial-sized AFBC boilers by requesting submittals of cost-sharing proposals for the following industrial categories:

Industry	SIC Code
Petroleum	29
Chemical	28
Primary metals	33
Paper and pulp	26
Food	20

If the potential for significant oil and gas savings is shown, the Program Opportunity Notice (PON) will invite industry proposals for four plants producing 90,800 kg/hr (200,000 lb/hr) steam.

2.2.2 State of Ohio's Development Program[25]

On other fronts, the State of Ohio is active in the commercialization of fluidized-bed combustion. During the natural gas shortage of the winter of 1976, it became clear to the state that coal must be used more widely than it had been. At the same time, the federal government was considering implementation of more stringent SO_2 emission standards. Since Ohio mines yield high sulfur coal, there was concern from the coal industry and the governor about possible loss of jobs and fulfillment of energy needs in the state. Therefore, a committee was established to investigate FBC as a possible answer to the problem. The committee's investigation led to plans for installation of three FBC boilers to demonstrate the feasibility of the technology as applied to Ohio's needs.

The Governor's Coal Use Committee selected Babcock Contractors, Inc. (a joint venture with Riley Stoker Corporation) to install a 27,000 kg/hr (60,000 lb/hr) steam retrofit FBC boiler at the Central Ohio Psychiatric Hospital. The unit will be used for space heating and will startup during 1980. The other two boilers are planned as new installations, one for space heating and process steam production, and the other for electricity generation. The former is a 45,000 kg/hr (100,000 lb/hr) steam unit planned for the Ohio State Penitentiary in Columbus. Design is progressing on the latter boiler which will be of utility size; 160,000 kg/hr (350,000 lb/hr) steam capacity to be installed at the Columbus and Southern Ohio Electric Company at Piqua, Ohio. Construction and start-up schedules for these two units are uncertain at this time.

2.2.3 Commercial Availability of Fluidized-Bed Boilers

Commercial orders for FBC boilers are progressing, and it appears that foreign boiler manufacturers have received a significant share of initial orders. This includes Babcock Contractors, Inc. with one boiler contracted in Ohio,[26] and Stone Platt of Netherton, England, having sold FBC boilers to Virginia Polytechnic Institute (an experimental unit) and General Motors.[27] These two boilers are currently scheduled for startup.

Johnston Boiler Company of Ferrysburg, Michigan claims four sales to date. These include an 18,160 kg/hr (40,000 lb/hr) steam coal-fired unit at the Central Soya Company in Ohio. Two wood-fired units have been sold, one of 9,080 kg/hr (20,000 lb/hr) steam capacity to the Herman Miller Company, a furniture manufacturer in Zeeland, Michigan, and a second of 4,540 kg/hr (10,000 lb/hr) steam capacity to the Pike Lumber Company in Atkron, Indiana. IBM, in Charlotte, North Carolina, purchased a 9,080 kg/hr (20,000 lb/hr) steam boiler capable of firing gas/oil with the potential to switch to coal. All of these units are scheduled for startup in late 1979 and 1980.[28]

FBC development is occurring internationally as shown in Section 2.2.5.1, Table 13, in the United Kingdom, West Germany, Canada, India, and other countries.

2.2.3.1 Users Satisfaction/Acceptance of First Generation FBC Boilers--

The demand for FBC industrial boilers will increase as:

- The reliability of FBC technology is commercially demonstrated through continuous boiler operation with effective emission control.
- The economics of FBC use are shown to be competitive with conventional systems controlled at similar efficiency for SO_2, NO_x, and particulate matter.
- Government regulations concerning energy policy evolve which emphasize coal use in new facilities.
- Environmental control requirements are more firmly defined.

The results of the ongoing DOE program, the Ohio program, and initial operating results with boilers sold by Johnston Boiler Company, Foster-Wheeler, Babcock Contractors, Inc., Stone Platt, and others will be of major importance in establishing demand for industrial FBC boilers in the future. Although bench-scale and pilot facilities have been operated, until actual commercial use for a year or more of continuous operation is demonstrated, widespread demand will not develop.

2.2.4 Summary of Existing Fluidized-Bed Units

Table 13 is a listing of industrial AFBC demonstration facilities and pilot-scale test facilities.

2.2.5 Applicability of Fluidized-Bed Combustion to Industrial Uses

2.2.5.1 Limitations by Boiler Type--

Fluidized-bed combustion can be used in place of practically any type of boiler (stoker, pulverized coal, gas/oil) in any application such as saturated/unsaturated steam, process heating (water, air, crude oil), and direct/indirect heating. FBC may also be used to advantage in instances where conventional technology is limited because of FBC's proven multifuel capability.

In the industrial boiler capacity size range of less than 73 MW_t (250 × 10^6 Btu/hr), it is expected that most, if not all FBC units, will operate at atmospheric pressure with a once-through sorbent processing scheme. Most industrial FBC boiler users probably will not have sufficient need for onsite electric power generation to justify the additional capital and operating costs and operational complexity associated with pressurized FBC systems. In addition, atmospheric systems are now commercially offered for industrial use. A similar argument of economics, operational complexity, and technological demonstration holds true for sorbent regeneration systems. It is expected that the normal industrial user will select a once-through sorbent operating scheme, due to its demonstrated simplicity and lower cost, at least in first generation FBC installations.

TABLE 13. AFBC COAL-FIRED DEMONSTRATION AND TEST UNITS

Developer	Capacity	Location	Sponsor	Status and comments
Industrial Demonstration Units				
Combustion Engineering	22,700 kg/hr (50,000 lb/hr) steam	Great Lakes Naval Training Base, Illinois	U.S. Department of Energy (Cosponsor)	Construction to begin during winter, 1978
Foster-Wheeler; Pope, Evans, and Robbins; and Georgetown University	45,400 kg/hr (100,000 lb/hr) steam	Georgetown University Washington, D.C.	U.S. Department of Energy (Cosponsor)	Startup scheduled for summer, 1979
Exxon Research and Engineering Co.	3 - 4 MW_t (10-15 × 10^6 Btu/hr)	Linden, New Jersey or Baytown, Texas	U.S. Department of Energy (Cosponsor)	This unit is a process crude oil heater and is currently in the pretesting and design evaluation phase
Battelle-Columbus Laboratories	11,350 kg/hr (25,000 lb/hr) steam	Columbus, Ohio	U.S. Department of Energy (Cosponsor)	Under design based on SSEUS test unit, see below
Foster-Wheeler; Pope, Evans, and Robbins	45,400 kg/hr (100,000 lb/hr) steam	Wilkes-Barre, Pennsylvania	U.S. Department of Energy	To be constructed under DOE anthracite culm program
Shamokin Area Industrial Corporation (SAIC); Curtiss-Wright, Dorr-Oliver; Stone and Webster	9,080 kg/hr 20,000 lb/hr) steam	Cellu Products (Paper Company) Shamokin, Pennsylvania	U.S. Department of Energy	To be constructed under DOE anthracite culm program
Fluidyne Engineering Co.; Deltrak; Nebraska Boiler Company	9,080 - 13,600 kg/hr (20,000 - 90,000 lb/hr) steam	GTE Sylvania, Towanda, Pennsylvania	U.S. Department of Energy	Replacement for existing oil-fired boiler, to be constructed under DOE anthracite culm program
Foster-Wheeler; Pope, Evans, and Robbins	88 MW_t (220 × 10^6 Btu/hr)	Rivesville, West Virginia	U.S. Department of Energy	Currently operating
Babcock and Wilcox, Ltd. (England)	12 MW_t (40 × 10^6 Btu/hr)	Renfrew, Scotland	-	Retrofit unit currently operating
Babcock and Wilcox Company (U.S.)	6 MW_t (20 × 10^6 Btu/hr)	Alliance, Ohio	Electric Power Research Institute	Currently operating
Morgantown Energy Research Center	18 MW_t (60 × 10^6 Btu/hr)	Morgantown, West Virginia	U.S. Department of Energy	Under design

(continued)

TABLE 13 (continued).

Developer	Capacity	Location	Sponsor	Status and comments
Babcock and Wilcox, Ltd. (England)	27,250 kg/hr (60,000 lb/hr) steam	Central Ohio Psychiatric Hospital Columbus, Ohio	Ohio Department of Energy	Startup scheduled for late 1979
To be negotiated	45,400 kg/hr (100,000 lb/hr) steam	Ohio State Penitentary Columbus, Ohio	Ohio Department of Energy	Retrofit installation currently in planning stage
To be negotiated	160,000 kg/hr (353,000 lb/hr) steam	Columbus and Southern Ohio Electric Company Piqua, Ohio	Ohio Department of Energy	Utility boiler currently in planning stage
Johnston Boiler Co.	3 MW_t (10×10^6 Btu/hr)	Johnston Boiler Co. Ferrysburg, Michigan	Private	Demonstration boiler currently operating.
Johnston Boiler Co.	18,200 kg/hr (40,000 lb/hr) steam	Central Soya	Private	Recently sold
Johnston Boiler Co.	9,080 kg/hr (20,000 lb/hr) steam	IBM, Charlotte, North Carolina	Private	Recently sold; designed as oil/gas unit capable of burning coal
Wormser Engineering, Inc.	6 MW_t (20×10^6 Btu/hr)	Lowell, Massachusetts	Private	Currently operating
Mustad and Son (Gjovik, Norway)	25 MW_t (85×10^6 Btu/hr)	Varmeverk (Heating Works) Enköping, Sweden	-	Startup currently scheduled
Coal Processing Consultants (B&W, Ltd.)	36,300 kg/hr (80,000 lb/hr) steam	Prince Edward Island	Private	Startup scheculed for 1982
Energy Equipment	13,600 kg/hr (30,000 lb/hr) steam	Cadbury SCHWPS Boarnville United Kingdom	Private	Currently operational
Ruhrkohle	35 MW_t (105×10^6 Btu/hr)	Düsseldorf-Flingern West Germany	Private	Startup scheduled for early 1979
Wesertal GMBH	125 MW_t (375×10^6 Btu/hr)	Hameln, West Germany	Private	Currently operational
Mitchel Engineering	36,300 kg/hr (80,000 lb/hr) steam	Don River, United Kingdom	British Steel	Currently operational

(continued)

TABLE 13 (continued).

Developer	Capacity	Location	Sponsor	Status and comments
Pilot Scale Test Units				
Combustion Engineering	1 MW_t (3×10^6 Btu/hr)	Windsor, Connecticut	U.S. Department of Energy	Currently operating
Energy Resources Company	1.8 MW_t (6×10^6 Btu/hr)	Cambridge, Massachusetts	Private	Currently operating
Pope, Evans, and Robbins	1.5 MW_t (5×10^6 Btu/hr)	Alexandria, Virginia	U.S. Department of Energy	Currently operating
Stone Platt Fluidfire, Ltd.	0.3 MW_t (1×10^6 Btu/hr)	Virginia Polytechnic Institute; Blackburg, Virginia	-	Startup currently scheduled
Stal-Laval Turbine Company (Finspaug, Sweden)	1.5 MW_t (4.5×10^6 Btu/hr)	District Heating Plant Orebro, Sweden	-	Currently operating
Fluidyne Engineering	≤5,700 kg/hr (12,600 lb/hr) hot air output	Minneapolis, Minnesota		
EPA Sampling and Analysis Test Rig (SATR)	≈0.3 MW_t (1×10^6 Btu/hr)	Research Triangle Park, North Carolina	U.S. Environmental Protection Agency	Currently operating
Babcock and Wilcox Company (U.S.)	1.5 MW_t (5×10^6 Btu/hr)	Alliance, Ohio	Electric Power Research Institute	Currently operating
Battelle-Columbus Laboratories (SSEUS)	1.5 MW_t (5×10^6 Btu/hr)	Columbus, Ohio	U.S. Department of Energy	Currently operating

Heat exchange media used in fluidized-bed boilers will include steam, air, and other fluids (e.g., process streams such as crude oil). In most units, heat transfer surface in the form of water or air tubes will be immersed directly in the fluidized bed to maximize heat transfer rate and efficiency. Convective transfer surfaces (water tube, fire tube, air tube) could be applied to act as superheater, preheater, or economizer.

2.2.5.2 Limitations by Fuel Characteristics--

Fuel flexibility is an important advantage of FBC use in the industrial sector due to the incentive to burn industrial byproducts and low-grade, high sulfur fuels not easily burned in conventional boilers. FBC boilers have multifuel capability and can burn all ranges of coal, oil, and gas and some industrial wastes.

Johnston Boiler is currently offering multifuel FBC boilers, having sold one coal-fired unit, one gas/oil unit (with coal-firing capability), and two wood-fired units. Other tests have been conducted with all types of coal including anthracite/anthracite culm at the Morgantown Energy Research Center and lignite at the Grand Forks Energy Research Center. Industrial byproduct waste combustion has also been demonstrated.

2.2.5.3 Limitations by Boiler Size--

The concensus of opinion indicates that widespread application of coal-fired FBC industrial boilers will be limited to systems greater than 15 to 30 MW_t (50 to 100 × 10^6 Btu/hr)[29-32] due primarily to the disporportionately high cost of related coal and ash handling equipment for smaller units. However, Johnston Boiler Company[33] is marketing coal-fired units as small as 1,140 kg/hr (2,400 lb/hr) steam which is roughly equal to 0.9 MW_t (3.1 × 10^6 Btu/hr). To date, the smallest unit they have sold expressly for coal-firing has a capacity of 18,160 kg/hr (40,000 lb/hr) steam or about 15 MW_t (50×10^6 Btu/hr). Johnston has also sold a gas/oil unit capable of coal-firing with a capacity of 4,540 kg/hr (10,000 lb/hr) steam or about 4 MW_t (13×10^6 Btu/hr). In general, if FBC industrial boilers are used in the size range <30 MW_t (<100×10^6 Btu/hr) they may be employed to burn oil or possibly gas with future conversion to coal based on trends in fuel availability and environmental standards.

The important feature of FBC with respect to boiler size is that it may extend to lower limits, the boiler size in which coal can be used due to lower system cost and the avoidance of SO_2 scrubbing. There does not appear to be any technical lower capacity limit to coal-firing with FBC technology.

FBC boilers have achieved heat release rates of >1 MW_t/m^3 (>100,000 $Btu/hr/ft^3$) of expanded bed volume or 0.5 to 0.6 MW_t/m^3 (50 to 60,000 Btu/hr/ ft^3) of firebox. This compares to a heat release rate of 0.2 MW_t/m^3 (20,000 $Btu/hr/ft^3$) of firebox in a conventional pulverized coal boiler.[34] Therefore, it is anticipated that package FBC units will be available in larger thermal capacities than conventional boilers.

First generation fluidized-bed combustion boilers will most likely be in the energy capacity range of less than 73 MW_t (250×10^6 Btu/hr) thermal input. Industrial, commercial and institutional facilities with new, additional or replacement energy needs will be the potential buyers for the FBC boilers in that category. Presently, there are over 3,000 United States boilers in this size category.[35]

The Fuels Use Act of 1978[36] may provide an incentive for use of coal-fired FBC boilers in capacities greater than 29 MW_t (100×10^6 Btu/hr). The legislation calls for use of coal-firing in all new boiler systems greater than this capacity unless the effectiveness of coal use can be proven unsuitable for technical or economic reasons.

A summary of expected FBC boiler configurations by size range is provided in Figure 3.

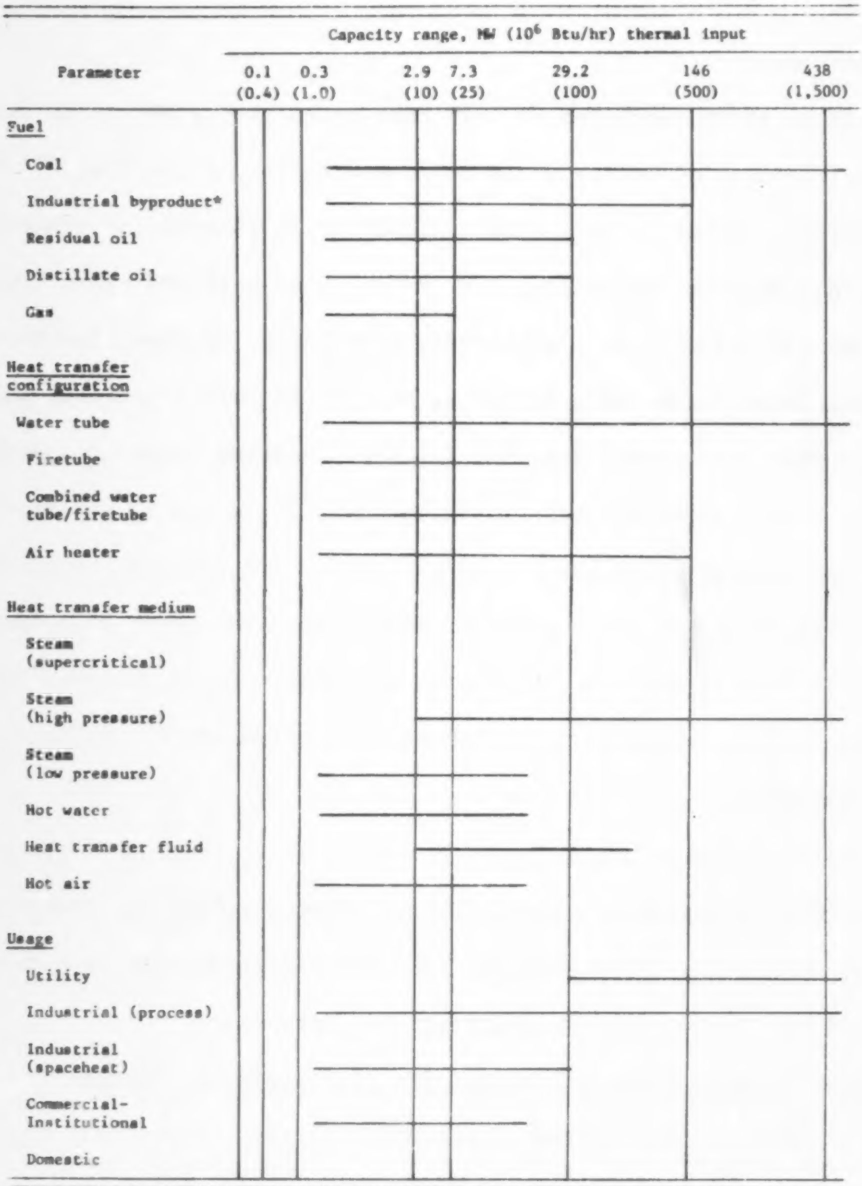
2.2.5.4 Retrofits--

A study by EXXON concluded in 1976 that retrofitting FBC to an existing conventional industrial boiler would be economically unattractive.[37] However, one retrofit FBC boiler is operating and another is planned for commercial installation. Babcock and Wilcox, Ltd. constructed a 18,000 kg/hr (40,000 lb/hr steam) FBC retrofit on a stoker-fired boiler in Renfrew, Scotland. They are planning installation of a 27,000 kg/hr (60,000 lb/hr) retrofit unit at the Central Ohio Psychiatric Hospital for space heating purposes. These retrofits are on stoker-fired boilers, where the existing grate is replaced with a fluidized bed incorporating heat exchange tubes. The existing convective heat transfer surfaces can be retained, thus minimizing the extent of conversion required. If retrofitting is considered, the stoker-fired boiler is the most appropriate system because actual conversion requirements are minimized and capacity downrating may not result.

The actual economic and technical feasibility of FBC retrofitting is not known, but will be extremely site-specific. However, based on these early ventures by B&W, Ltd., it is apparent that FBC technology can be considered in instances where system retrofitting might be appropriate.

2.2.6 Projections of Potential Market for Fluidized-Bed Combustion

Farmer, et al., have estimated potential national industrial FBC boiler application through the year 2000.[38] Most of the potential is expected to be in the chemicals, petrochemicals, petroleum refining, paper, primary metals, and food industries which are the industrial categories with the heaviest steam demand. These projections were made in 1976. Since the current



*May include low grade low cost fuels such as lignite, bark and wood waste, process tars, and sludges.

Figure 3. Atmospheric FBC industrial boilers — occurrence of various boiler parameters by capacity range.

industrial boiler market is depressed in general, the forecast may be high. GCA's independent investigation indicates that current FBC vendors have the capability to fabricate the number of boilers indicated. However, the demand is uncertain. The nationwide potential was projected as follows:

Year	Cumulative number of industrial FBC boilers	10^{15} Btu per year	1,000 B/D of oil equivalent
1980	7	0.01	5
1985	200	0.29	136
1990	685	0.99	462
1995	1170	1.69	793
2000	2050	2.97	1400

2.2.7 Recent Improvements and Ongoing Research and Development

2.2.7.1 Sulfur Dioxide Control--

Careful design of gas phase residence time and sorbent particle size can result in efficient SO_2 removal according to current projections by Westinghouse.[39] Model development by Westinghouse and others is continuing in order to model sulfur retention as influenced by these design and operating parameters.

The emphasis of future research will be confirmation of SO_2 control estimates in large-scale units. Documentation of the influence of gas phase residence time and sorbent particle size in large demonstration units is of prominent importance. The trade-offs associated with maximizing or minimizing these parameters must be defined.

Other investigations are required to assess limestone characteristics and availability as well as alternative sorbents. Energy Resources Company (ERCO) has recently begun investigation of interquarry limestone characteristics.[40] This study should give a good perspective of the effects of limestone variations. Westinghouse will be conducting a detailed investigation of intraquarry variations.[41]

The Illinois State Geological Survey has extensively studied several varieties of carbonate rock (mainly limestone and dolomite) for desulfurization in fossil fuel combustion processes.[42] Samples were investigated for petrography, mineralogy, chemistry, pore structure, and surface area. A wide range of petrographic and SO_2 sorptive properties were revealed. Relatively high SO_2 reactivity was found for chalks, calcareous marls, and oolitic aragonite sand samples, probably due to high pore volumes and fine grain size.

General Electric is conducting experimentation to develop an automatic process controller to maintain a constant percentage of SO_2 removal by the bed.[43] This capability is necessary to adjust for changing bed conditions without allowing excessive SO_2 emissions for intermittent periods. Expanded research and development in the area is expected.

Experimentation with additives for improved desulfurization has been conducted. Argonne National Laboratories has studied the effect of adding NaCl to the bed.[44] Although the pore surface area and calcium utilization are increased by salt addition, salt has a great potential for producing boiler corrosion. Other catalysts under consideration are iron oxide and coal ash.

Westinghouse[45] has done some preliminary investigations of Na_2CO_3, $NaAlO_2$, $NaCO_3$, Fe_2O_3, and $CaAl_2O_4$ as alternative sorbents. Investigators at Argonne National Laboratories are experimenting with virgin and spent oil shale.[46] Virgin shale is attractive because of its inherent heating value of about 3,000 Btu/lb.

Sorbent regeneration techniques also require further exploration and development to minimize feed requirements, spent stone disposal, and associated sensible heat loss. EXXON is attempting to develop regenerable synthetic sorbents that have good attrition resistance, high reactivity, and good regeneration characteristics.[47] Calcium aluminate cement and calcium or barium titanate both appear to have characteristics which may make these materials cost competitive with limestone. Methods of enhancing limestone reactivity by precalcining (currently under investigation at EXXON[48]) and catalyst addition must also be studied.

In essence, the thrust of current and future work is the minimization of sorbent requirements and spent stone disposal to optimize SO_2 retention and minimize cost, energy, and environmental impact.

2.2.7.2 Nitrogen Oxides Control--

The emphasis of past research has been to document emissions from experimental AFBC units being operated for some experimental purpose other than deliberate NO_x control. Little has been done to reduce NO emissions (generally between 129 to 258 ng/J (0.3 to 0.6 $lb/10^6$ Btu)[49]) measured during normal operation at FBC test units, other than to generally observe the impact on emissions as experimental conditions were being varied for some other purpose. Experimental and modeling work is continuing in an effort to gain a better understanding of NO_x formation/reduction mechanisms in FBC, and of the correlation between emissions and the key FBC design/operating conditions which can influence emissions. The goal of these studies is to provide the capability to better predict and control NO_x emissions through simple adjustment of standard design/operating conditions. Also, several investigators are beginning to address combustion modifications, deliberately aimed at reducing NO_x emissions from FBC, such as staged combustion, flue gas recirculation, ammonia/urea injection, and stacked beds. It is necessary to define the effects of such combustion modification techniques, not only on NO_x emissions, but on other system parameters, such as combustion efficiency and materials corrosion and the potential increase of SO_2 or particulate emissions.

2.2.7.3 Particulate Control--

The major requirement in this area is to test conventional particulate control devices applied to AFBC boilers. Although performance is not documented, it should be similar to conventional systems burning low sulfur coal. Testing is currently being performed at the Sampling and Analytic Test Rig (SATR) operated by the U.S. Environmental Protection Agency.[50] Testing is also planned at the 30 MW_e (300,000 lb/hr steam) demonstration facility in Rivesville, West Virginia, the 10 MW_t (100,000 lb/hr steam output) unit under construction at Georgetown University, and other FBC units as they become available.

2.2.7.4 Solid Residue Disposal/Utilization

The disposal and utilization character of FBC solid waste should be the focus of considerable investigation in the near future. It is imperative that optional disposal and handling methods are assessed and ways to minimize the environmental, cost and energy impact of disposal are found, due to the large volume of material which will be produced as commercial units are brought online.

The waste may be usable for commercial purposes. Presently two main areas are under investigation, use as a structural material like concrete or use as an agricultural soil conditioner.

Several studies have demonstrated that FBC solid residue are cementitious. This characteristic can be exploited to form a very durable concrete-like mass. One DOE study is under way to investigate the potential of using FBC solid waste for road construction.[51] The results indicated that compressive strength of cemented waste exceeded the value recommended for heavy traffic highway construction over a wide range of compositions. Further, this compressive strength, which is indicative of the material durability and resistance to erosion, improved with time even after the cemented samples were subjected to freeze/thaw cycles. The study concluded that the exceptional high strength of cemented FBC residue makes it suitable for applications which require materials with low water permeability, such as in embankment, structural fill, and liners to control leaching from waste disposal landfills and lagoons.

Another DOE study being performed simultaneously in several states in the eastern United States is an agricultural application study for FBC solid waste.[52] The program covers almost all the varieties of crops grown in the eastern United States. It includes both short- and long-term laboratory and field-based evaluations. The waste is used as a replacement for lime to neutralize soil, as a source for trace and certain nutrient elements, and as a source for sulfur. The study evaluates both the quality and quantity of crops produced from soil treated by waste material, as well as the crops' nutrient value as food for domestic animals.

A study to evaluate the physiological effects of food that is ultimately obtained from FBC waste-treated soils on people and animals has been proposed to DOE and EPA. The study will monitor mineral balance and amino acids in human tissues, primarily human hairs, which tend to accumulate toxic materials. Some small animals will be evaluated over several reproductive cycles to determine long-term effects on offspring. The first stage of tests will start in October 1979 and the second stage is scheduled for 1980.

Further investigation of uses for solid waste from FBC are necessary. By finding viable commercial uses for the residue, the environmental and cost impact of FBC would be greatly reduced.

2.2.7.5 Other Investigations--

The performance capability and cost of inbed versus overbed solids feeding is an important issue under study. Although several developer/vendors are engineering systems using either of these techniques, the most current study of the effect of feed orientation on SO_2 control is being conducted by FluiDyne (in their 3.3 ft × 5.3 ft unit) for EPA.[53] This study will assess performance as a function of feed orientation, gas residence time, limestone particle size, and use of primary recycle. Earlier experiments by FluiDyne in their 18 in. × 18 in. unit indicated that equivalent desulfurization could be achieved regardless of feed orientation as long as primary recycle was practiced (see Section 7.0).[54]

2.3 SYSTEM PERFORMANCE - SO_2 CONTROL

This and the following two sections describe the key variables affecting the performance of FBC units in terms of emissions of SO_2, NO_x and particulates. In the absence of data from large FBC facilities, much of this discussion is based upon data from experimental units, and the results from modeling activities. Data from large facilities are necessary to confirm the absolute performance that will be observed in commercial FBC installations.

One of the major advantages of FBC over conventional combustion of coal is that SO_2 is removed within the bed using a calcium-based sorbent. Design operating factors which influence the control of SO_2 emissions for an atmospheric fluidized-bed combustor (AFBC) follow:

- Primary factors - Ca/S molar feed ratio
  - sorbent particle size
  - gas phase residence time

$$\left(\begin{matrix}\text{expanded}\\ \text{bed height}\end{matrix} \div \begin{matrix}\text{superficial}\\ \text{gas velocity}\end{matrix}\right)$$

- Secondary factors - sorbent reactivity
  - bed temperature
  - feed mechanisms
  - excess air

2.3.1 Primary Design/Operating Factors Affecting SO_2 Emission Reduction

SO_2 produced during the combustion of coal is reduced in FBC by burning the fuel in the presence of calcium oxide. The SO_2 reacts with the calcium oxide and excess oxygen forming calcium sulfate.

$$SO_2 + CaO + 1/2O_2 \rightarrow CaSO_4 \text{ (anhydrous)}$$

Calcium-based sorbents such as lime, limestone and dolomite are the most commonly used sorbents for FBC. The calcium content is the constituent which determines the amount of sorbent required to reduce the SO_2 emissions to a given level. (Availability of the calcium for reaction depends on sorbent type, particle size, gas phase residence time, and the extent of sulfation.) Thus the ratio of the calcium content of the sorbent to the sulfur content of the coal is used to determine sorbent needs to control SO_2.

2.3.1.1 Ca/S Ratio--

Of the factors which affect SO_2 emission control, the calcium to sulfur molar feed ratio (Ca/S) has the greatest impact. As the calcium content of the bed is increased, greater SO_2 removal is achieved. Westinghouse Research and Development Center has developed a model which projects sorbent requirements to attain certain levels of SO_2 removal efficiency. Figure 4 illustrates the rapid increase in sulfur retention with increasing Ca/S based on the model.[55] For sorbents with a particle size of approximately 500 μm, the relationship is nearly linear below about 75 percent SO_2 removal. Above this level, sulfur retention approaches 100 percent asymptomatically. Experimental test data, where available, concur with the projections (see Section 7.0). However, further data from larger systems and for high levels of SO_2 removal are required to support the model projections. The Westinghouse desulfurization model assumes uniform sulfur generation throughout the bed. In underbed feed systems where SO_2 may be preferentially formed near the bottom of the bed, the Westinghouse model may underpredict the SO_2 reduction capability of the FBC system.

The curves shown in Figure 4 for Greer, Grove, and Carbon Limestone are taken from a recent Westinghouse report.[56] Westinghouse is currently investigating industrial FBC boilers in their study "Effect of SO_2 Emission Requirements on Fluidized-Bed Boilers for Industrial Applications: Preliminary Technical/Economic Assessment."[57] The Western, Bussen, and Menlo quarry limestones shown in Figure 4 are the basic sorbents used in their industrial boiler study as examples of high, medium, and low reactivity sorbents, respectively. The least reactive sorbent (Menlo) or one with similarily low reactivity would probably be avoided in practice because a Ca/S ratio close to six is required to achieve 90 percent SO_2 removal (under "best system" conditions, as discussed in Section 3.0). Better sorbent should be routinely available to industrial customers.

The data shown are based on an average inbed surface particle diameter of 500 μm, and the assumption that primary particle recirculation will be used. Primary recycle should prove cost effective from the standpoint of improved SO_2 control and combustion efficiency. If primary recycle were not used, a

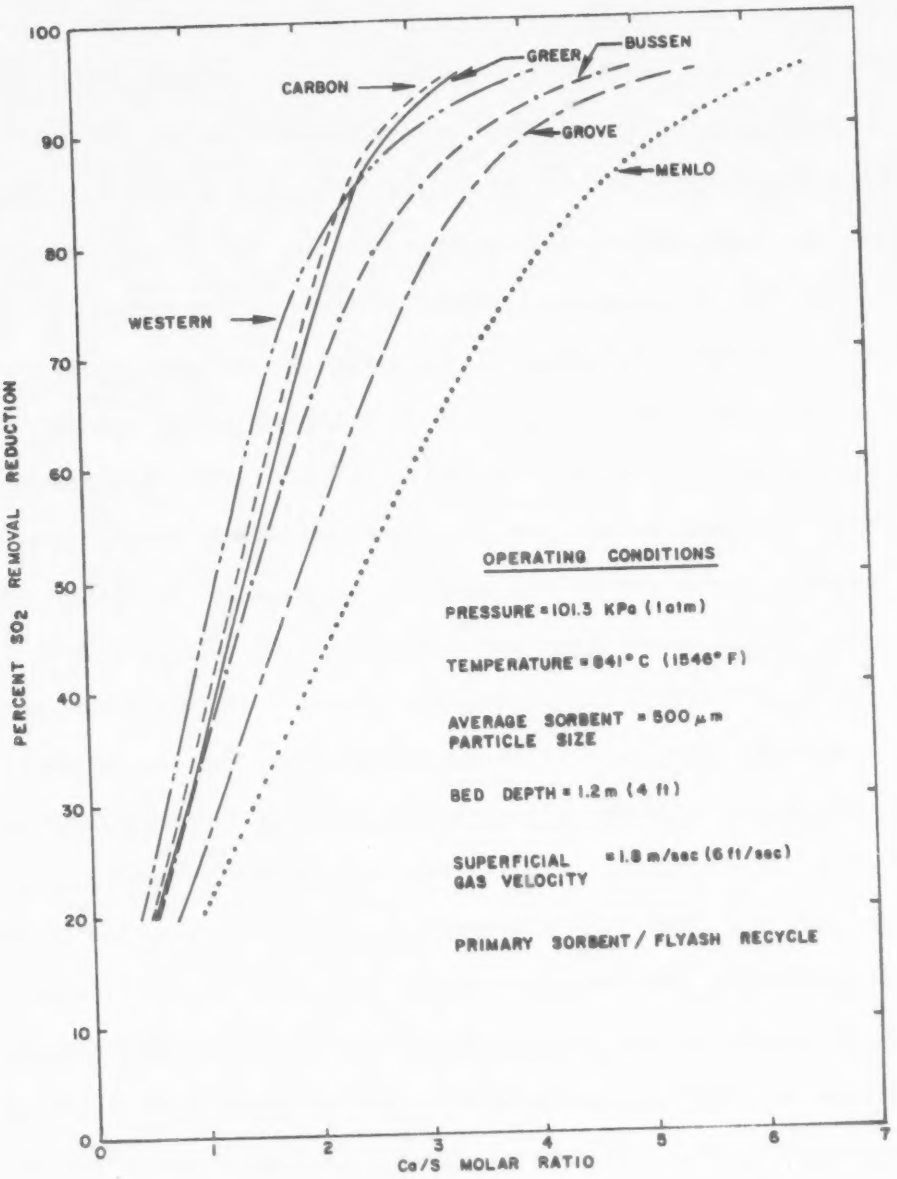


Figure 4. Projected desulfurization performance of atmospheric fluidized-bed coal combustor, based upon model developed by Westinghouse.[55]

coarser sorbent might be required (inbed average of 1,000 μm or greater) to avoid unacceptable sorbent losses, and Ca/S molar feed requirements would increase substantially. (As discussed in Section 3.0, primary recycle is considered an important feature of "best system" design for SO_2 control.)

Table 14 summarizes some of the available data on sulfur retention versus Ca/S molar feed ratio and sorbent particle size for several limestones. Again, the Ca/S ratio must be increased to achieve higher sulfur removal efficiency. Although total sorbent quantities will be different the same sulfur removal efficiency can be achieved burning coals of different sulfur concentration by maintaining the same Ca/S molar feed ratio, if all of the other key operating/design conditions (such as gas residence time) are maintained the same, and as long as the first order sorbent/SO_2 reaction kinetics do not change. For low sulfur coals, the reaction mechanism could conceivably change at very low SO_2 partial pressures. Under these conditions, if the coal sulfur concentration increases, the same level of control can be maintained by increasing the calcium feed proportionally. Figure 5 illustrates this using limestone 1359 to reduce emissions from the combustion of coals with 2.6 and 4.5 percent sulfur.[58] These tests were run under the same conditions with the exception of the difference in the coal sulfur content. Notice that the sulfur retention versus Ca/S ratio is better in this experimental case than in the Westinghouse projection in Figure 4. This may be due to the finer particle size of the sorbent in the experimental case. As Table 14 clearly indicates, the particle size of the sorbent is a major factor in SO_2 capture.

2.3.1.2 Limestone Particle Size--

As the particle size of a given sorbent is decreased, the calcium utilization is increased. Thus, with the same Ca/S molar feed ratio, the SO_2 reduction

TABLE 14. AFBC — Ca/S MOLAR FEED RATIOS OBSERVED TO MEET STRINGENT, INTERMEDIATE, AND MODERATE SO_2 EMISSION CONTROL LEVELS

Sorbent		Ca/S molar feed ratios required to meet optional control levels			References		
Type	Particle size (μm)	Stringent 90%	Intermediate 85%	Moderate 75%	Organization	Unit ID	Number
Limestone 1359	420 - 500	3.9	3.5	2.8	Westinghouse	*	45
	490 - 630	3.5	3.0	2.2	Argonne	6" diam.	46
	630	3.5	-	-	Argonne	6" diam.	47
	930	5.5	-	-	Exxon	3" diam.	48
	1,000 - 2,380	6.0	5.7	4.6	Babcock & Wilcox	3' × 3'	49
Greer Limestone	420 - 500	2.8	2.6	2.2	Westinghouse	*	45
	1,000 - 2,380	4.5	4.2	3.5	Babcock & Wilcox	3' × 3'	49
Carbon Limestone	420 - 500	2.6	2.4	2.0	Westinghouse	*	45
	500	2.9	-	-	Westinghouse	*	50
	1,000	7.0	-	-	Westinghouse	*	50
Limestone 1360	630	-	-	2.3	Argonne	6" diam.	51
	1,000 - 1,400	-	4.2	2.6	Argonne	6" diam.	52
Limestone 18	<1,680 453 median	4.0	3.6	3.1	National Coal Board of England	CRE	53
	<3,175	5.2	4.8	4.1	National Coal Board of England	CRE	53
Lowellville Limestone	1,000 - 2,380	5.5	5.0	4.0	Babcock & Wilcox	3' × 3'	49
Tymochtee Dolomite	630	2.6	-	-	Argonne	6" diam.	47
Hydrated Lime	<44	3.0	2.8	2.1	Babcock & Wilcox	3' × 3'	49
Western 90% CaL	500	2.8	2.5	1.9	Westinghouse	*	54
Bussen Quarry	500	3.4	2.9	2.3	Westinghouse	*	54
Menlo Quarry	500	5.3	4.7	3.9	Westinghouse	*	54

*These data points are based on the Westinghouse model; all others are experimental data.

Note: Temp - 540° to 980°C (1,000° to 1,800°F);
Excess air - 18 to 20 percent.

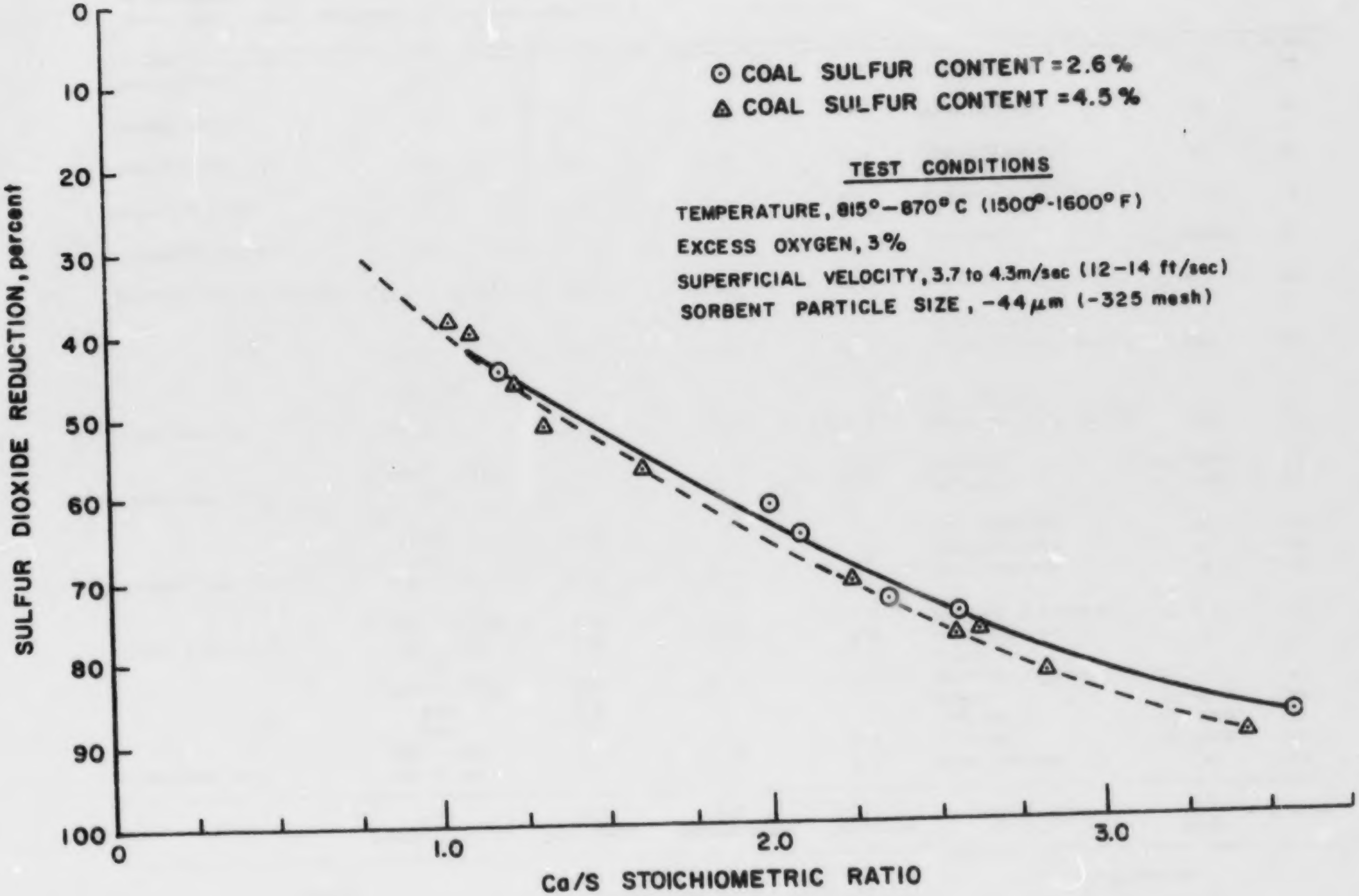


Figure 5. Sulfur dioxide reduction using limestone 1359 in a bed of sintered ash, Pope, Evans, and Robbins.[44]

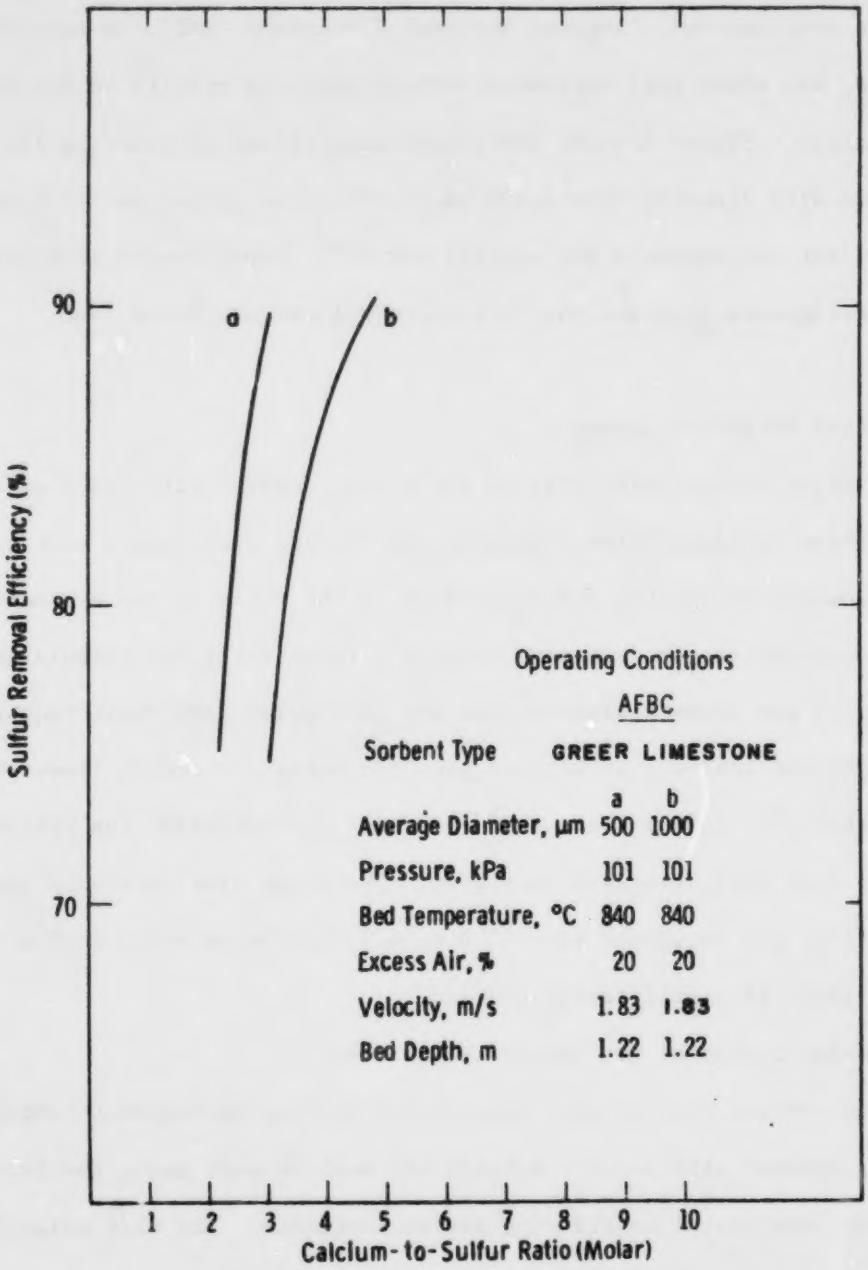
efficiency can be increased significantly by decreasing the sorbent particle size. The increased reactivity of smaller sorbent particles is due to the greater surface area exposed. Argonne National Laboratory (ANL), in controlled sorbent studies, has shown that increased sorbent porosity results in increased calcium utilization. Figure 6 shows the significant effect of reducing the average particle size diameter from 1,000 μm to 500 μm as projected for Greer limestone using the Westinghouse SO_2 kinetic model.[59] Experimental test data by several investigators indicate that these projections are valid (see Section 7.0).

2.3.1.3 Gas Phase Residence Time--

The third major factor which affects the sulfur removal efficiency of the system is gas phase residence time. This is the average time period that a unit volume of gas remains in the bed and is defined as the ratio of the expanded bed height to the superficial gas velocity. Figure 7 illustrates the calculated relationship between gas phase residence time and Ca/S molar feed ratio required to achieve 90 percent control, at various particle sizes for Carbon limestone and Grove limestone.[60] As gas phase residence time is increased, the calcium to sulfur molar feed ratio required decreases. The graph also indicates that there is a critical gas residence time (0.6 to 0.7 sec) below which sulfur retention efficiency is significantly reduced.

2.3.1.4 Interrelationship of Key Control Variables--

These three control factors are interrelated and can be varied to obtain the optimum SO_2 removal efficiency. A trade-off must be made among the factors to ensure the optimum system considering system economics. The Ca/S molar feed ratio required for a given level of control can be reduced by decreasing particle size or increasing gas residence time. However, if the particle size is



Operating Conditions

AFBC

Sorbent Type	GREER LIMESTONE	
	a	b
Average Diameter, µm	500	1000
Pressure, kPa	101	101
Bed Temperature, °C	840	840
Excess Air, %	20	20
Velocity, m/s	1.83	1.83
Bed Depth, m	1.22	1.22

Figure 6. Sulfur removal performance for typical sorbents (projected using Westinghouse kinetic model).

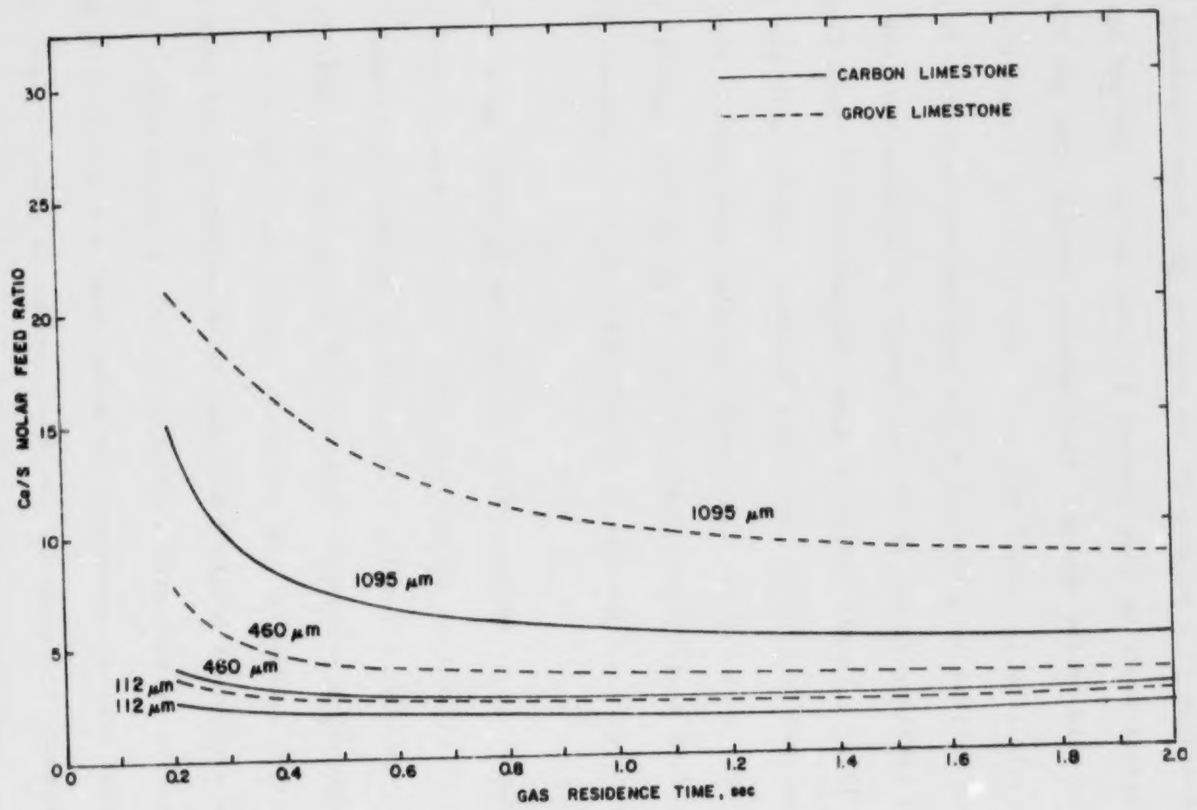


Figure 7. Ca/S molar feed required to maintain 90 percent sulfur removal in AFBC, as projected by the Westinghouse Model.[60]

decreased the gas velocity must be decreased so that the particles will not elutriate from the bed. This in turn increases the gas phase residence time. The optimum system is a balance of the minimum gas phase residence time which gives sufficient reaction time (around 0.7 sec) and the minimum particle size which can be used in the system. Westinghouse results indicate that an appropriate particle size is around 500 µm.[61] Figures 8 and 9 show the relationship of the three factors as predicted by the Westinghouse Model for 90 percent sulfur removal considering one of the more reactive (carbon) and less reactive (limestone 1359) sorbents tested to date, respectively.[62] Both figures show that the required Ca/S molar feed ratio increases rapidly with gas phase residence time less than 0.8 sec and sorbent particle size greater than 700. Under these conditions Westinghouse predicts that 90 percent SO_2 removal can be achieved using Carbon limestone at a Ca/S ratio of 3 or limestone 1359 at a Ca/S ratio of 5.

In summary, it is apparent that the calcium to sulfur molar feed ratio, the sorbent particle size and the gas phase residence time provide the key to the best SO_2 emission reduction performance in fluidized-bed combustion.

To increase gas residence times to 0.67 sec or greater (most "commercially-offered" designs operate at gas residence time in the range of 0.4 to 0.5 sec), boiler cross section or height would have to be expanded. The cost impact of this modification is discussed in Section 4.3.4. Although boiler expansion requires higher capital investment for added steel and potentially greater coal feeding equipment, there may be resultant savings in other capital equipment costs such as particulate control equipment (due to lower elutriation) or recirculation pumps (if natural circulation can be achieved using deeper beds).

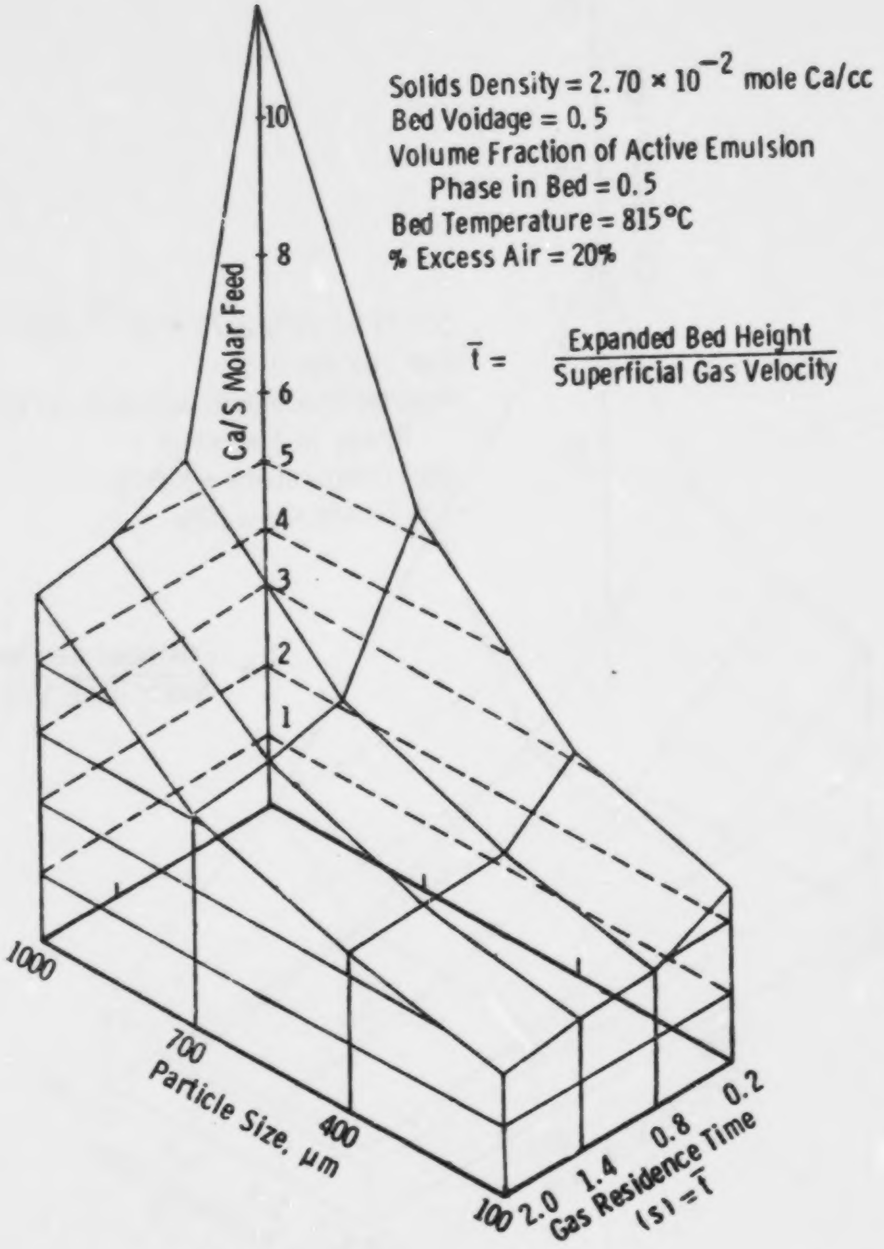


Figure 8. Ca/S molar feed required to maintain 90 percent sulfur removal in AFBC with Carbon limestone, as projected by the Westinghouse Model.[62]

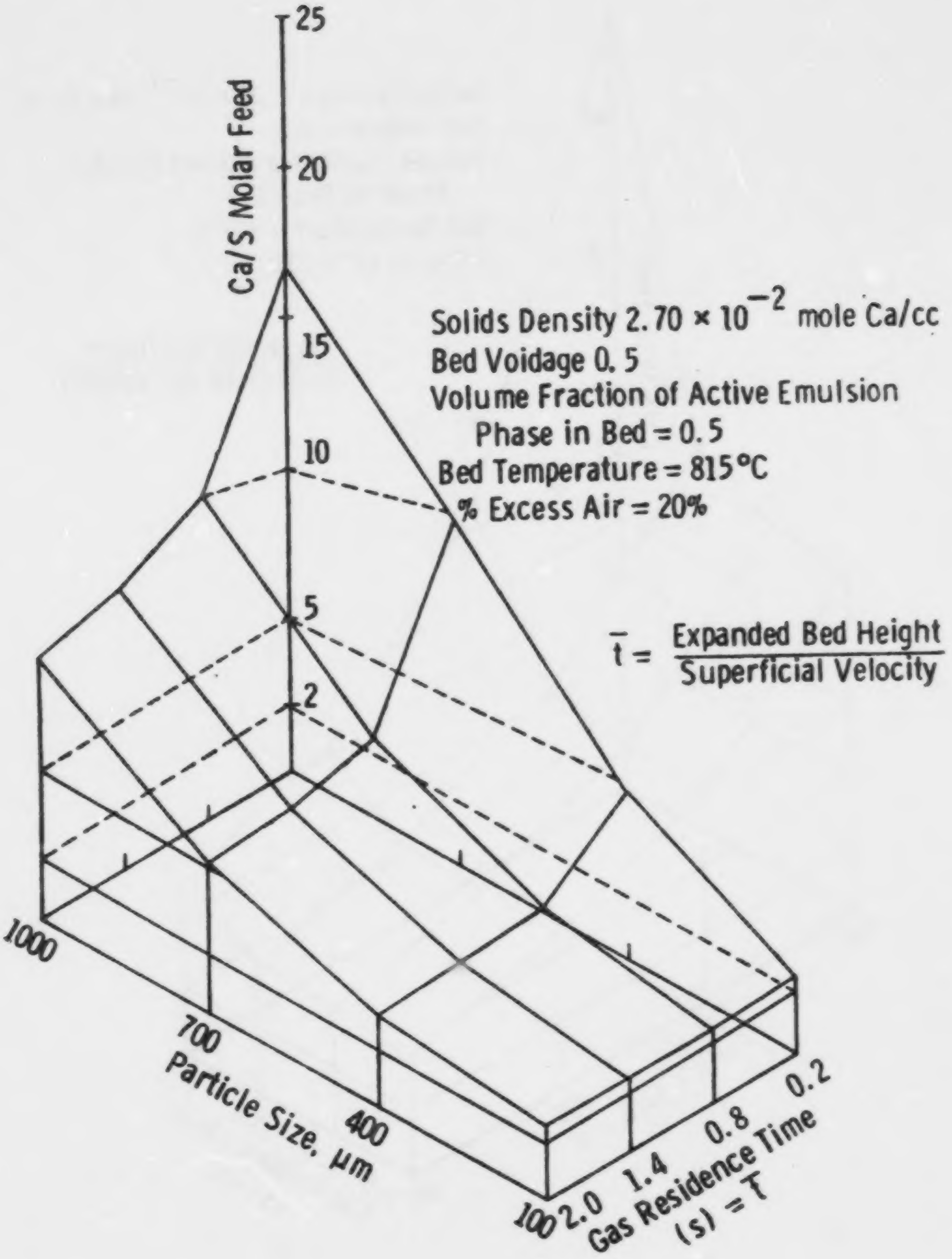


Figure 9. Ca/S molar feed required to maintain 90 percent sulfur removal in AFBC with limestone 1359, as projected by the Westinghouse Model.[62]

It is expected that these capital savings and more importantly operating savings in sorbent use, electricity, and improved combustion efficiency may offset the added capital cost associated with lengthening gas residence time.

2.3.1.5 SO_2 Emission Data Summary--

Figure 10 is a summary of SO_2 data obtained at eight AFBC test facilities under a wide variety of test conditions. The bounded area is an indication of the range of performance expected from FBC systems at high gas phase residence times and small sorbent particle size. Much of the experimental data falls within these boundries. The major excursions from the band are noted in the data from the B&W 3 ft × 3 ft unit and the PER-FBM unit. If the units and test conditions are considered closely (see Section 7.0) these deviations from the band are expected. The B&W 3 ft × 3 ft unit has a shallow bed which allows less than optimum sorbent/gas contact. Gas phase residence times are approximately one-third of 0.67 sec which is suggested for good reaction time. The PER-FBM data were also obtained using low gas phase residence times, in the range of 0.13 to 0.26 sec.

2.3.2 Secondary Factors Affecting SO_2 Reduction

The other factors which affect the performance of the SO_2 removal system are secondary, but can be used to obtain the maximum efficiency. Sorbent characteristics directly affect the Ca/S molar feed ratio. The temperature, solids feed mechanism, and excess air affect the rate and efficiency of the reaction between available CaO and SO_2.

2.3.2.1 Sorbent Characteristics--

The chemical and physical properties of a sorbent (i.e., sorbent reactivity) provide a basis for determination of sorbent requirements for a given combustion system. The volume of sorbent which will provide the desired sulfur retention

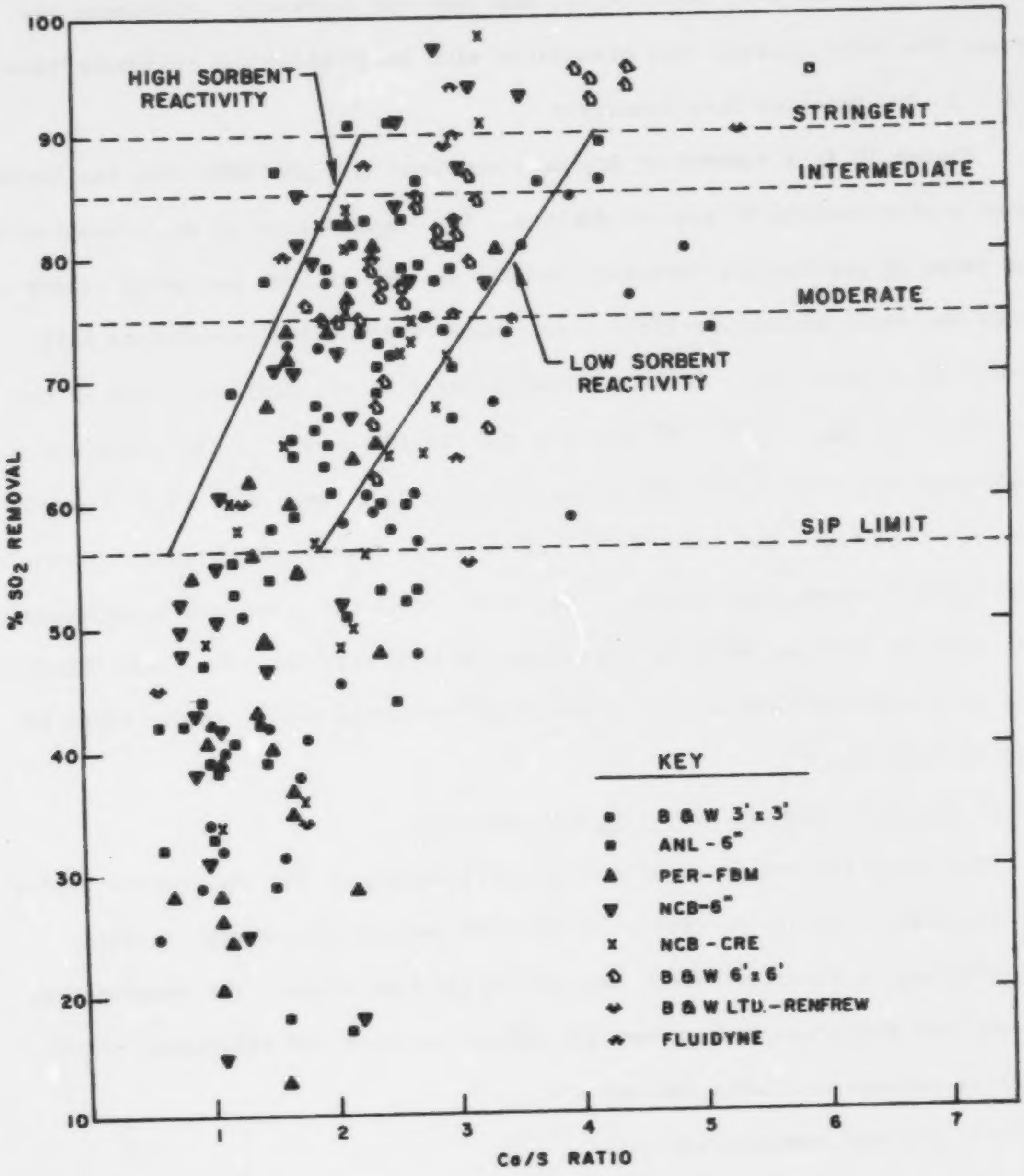


Figure 10. Summary of experimental SO_2 reduction data for AFBC test units.

will vary according to the calcium availability of the sorbent as well as its calcium content. Sorbent characterization and development studies by several investigators have identified the following factors which affect the sorbent reactivity:

- Sorbent porosity is the key to calcium utilization. Greater porosity increases the amount of surface area available for the gas/solid reaction.
- Sorbents which contain $MgCO_3$ have a slightly different grain structure than $CaCO_3$ alone. This grain structure provides greater pore surface area and thus greater calcium utilization potential.
- Dolomite, due to its $MgCO_3$ content, usually will have a better calcium utilization rate than limestone. However, a greater volume of dolomite is needed to obtain the same Ca/S ratio and thus equal or more solid waste may be generated.

Argonne National Laboratories performed thermogravimetric testing on 61 limestones for reactivity with SO_2 at 900°C using a gas mixture containing 0.3 percent SO_2.[63] There is large variability in the SO_2 reactivity of limestones and in the extent of conversion of the calcium carbonate to calcium sulfate. For the high calcium (>90 percent $CaCO_3$) limestones tested, the conversion of $CaCO_3$ to $CaSO_4$ ranged from 19 to 66 percent; for the dolomites (40 to 60 percent $CaCO_3$), the range was 21 to 100 percent.

Limestone availability is also an important factor in the development of FBC. The U.S. Environmental Protection Agency is initiating a broad sorbent screening study covering interquarry and intraquarry characterization. Although there appears to be no forseeable problem in sorbent availability, the quality of the material may have an impact on the FBC sorbent market. Limestone for fluidized-bed combustion must not only have good chemical reactivity but must meet physical standards for specific gravity, bulk density, crushing strength, loss of abrasion, porosity and toughness. The major requirement for the commercial use of limestone is the particle size of the rock. In the mining and preparation of the stone, a considerable amount of off-size material is produced. This material is stockpiled for use in other commercial uses; not all the limestone mined can be used for FBC.[64]

2.3.2.2 Temperature--

The temperature within the bed may have a direct effect on the efficiency of the reaction between sulfur dioxide and calcium oxide. Several investigators have shown that the optimum temperature for calcium use is between 760° and 870°C (1400° to 1600°F), depending upon the coal and sorbent in use.[65] Figure 11 shows the results of a study by Argonne National Laboratory on a 6-inch diameter AFBC system.[66] The temperature lower limit is determined by the temperature at which calcination occurs; that is, $CaCO_3$ releases CO_2, forming CaO, the reactive form of the sorbent. Below 760°C (1400°F) calcination is not complete. The lower sulfur retention above the optimum temperature may be caused by the release of SO_2 after capture due to local reducing conditions in the bed,[67] or by slight changes in other operating variables.

Experimental data have shown that within the bed there are oxidizing and reducing zones which affect the reactivity of the sorbent. Sorbent particles which migrate between the zones will produce greater sulfur capture than particles which are exposed only to a reducing environment.

2.3.2.3 Feed Mechanisms--

The sorbent and coal feed points can also affect the calcium utilization rate. The boiler can be fed either from over the bed or under the bed. Generally underbed feed produces greater turbulence, allowing the sorbent particles to travel freely between oxidizing and reducing zones. However, overbed feed systems have also been shown to achieve good calcium utilization as long as the elutriated fines are recycled.

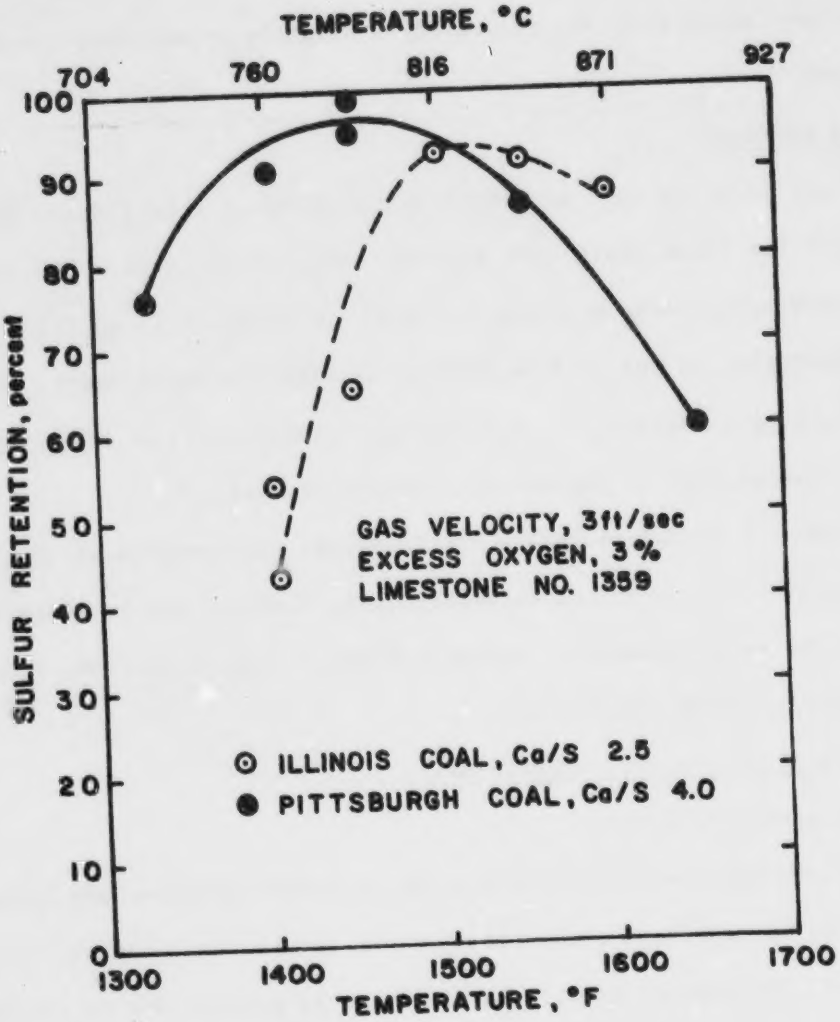


Figure 11. SO_2 reduction as a function of bed temperature (ANL).[66]

2.3.2.4 Excess Air--

The excess oxygen level has a lesser, but real effect on SO_2 capture. Investigators have found that SO_2 reduction is slightly enhanced by the increased excess oxygen.[68]

2.3.3 Other Factors

Variations from the mode of operation discussed in this section will not be significant for first generation boilers. However, in future FBC applications, new developing techniques may be used. Development is anticipated in new sorbent technologies in the form of sorbent utilization enhancement, regeneration and alternative sorbents, as well as technologies such as "fast" and "turbulent" fluidization to improve the combustion system.

Development is needed to reduce the limestone requirements so that the impact on limestone requirements and solid waste disposal can be minimized. In addition, the establishment of suitable modes of transportation, storage, and dust control must be considered.

2.3.4 Factors Affecting Boiler Performance

2.3.4.1 Corrosion/Erosion--

In fluidized-bed combustion boilers the corrosion problems are likely to be less than in conventional combustion boilers due to the lower bed temperature. However, the wear by erosion is likely to be greater due to the impact of the particles against boiler tubes and walls.

The erosion of heat transfer tubes within the bed is affected by the following factors:

- Coal particle size
- Sorbent particle size
- Chemical catalysts
- Bed temperature

- Particle velocity
- Oxidizing/reducing atmosphere

Larger coal and sorbent particle sizes produce greater potential tube erosion.[69] Smaller particles tend to follow the air stream around the tubes so that particles either fail to impact or do so at a lesser angle. The amount of erosion which a particle can produce is directly proportional to the angle of impingement of the particle.[70] The velocity, hardness, and sharpness of the particle can also be directly correlated with the degree of wear.[71] Vertical tubes would eliminate some of these effects, but the elbows or turns would still be highly susceptible to erosion. In addition, if chemical or thermal corrosion or degradation of material occurs, it will increase the affect of the erosion and abrasion. Temperatures within the range of FBC operating conditions seem to have little affect on the wear characteristics of the boiler.

The addition of NaCl, as proposed by some early researchers, to enhance calcium utilization may cause chemical corrosion in the form of pitting due to the reaction of the salt with the protective metal oxide coating on the tubes.[72] Generally pitting problems are not unique to FBC and can be controlled. However, whether salts are added to enhance Ca utilization or not, the migration of oxidizing and reducing zones within a turbulent bed (e.g., as a bubble moves up through the bed and around the immersed tubes) may have a detrimental effect on superheater tubes immersed in the bed at temperatures greater than 370°C (700°F).[73] Most tests have been conducted with metal temperatures less than 230°C (450°F). Further study of this phenomenon at higher metal temperatures is needed.*

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*The use of $CaCl_2$ may be better than NaCl and studies of this sort are also underway.

Generally it can be stated that if the proper materials are chosen for boiler walls and tubes, there should be little problem with erosion. The erosion properties of the construction material are inversely proportional to the surface hardness of an annealed material.[74]

2.3.4.2 Reliability and Turndown Capability--

The reliability of FBC has not yet been proven. Demonstration units are presently in the early phases of operation at best. The reliability of the systems will be better assessed within the next year.

Turndown in AFBC can be achieved in two ways: (1) by slumping one or more of several modules of the boiler; or (2) by reducing the bed depth of all the modules. The former is preferred because it is easier to maintain high sulfur capture. If the bed depth of all the modules is lowered the gas phase residence time will be reduced, and thus sulfur capture efficiency will decrease. With this in mind the turndown rate capability of AFBC could be dependent upon the number of cells which make up the boiler system.

2.3.4.3 Monitoring Needs--

The only additional monitoring need unique to FBC systems as opposed to conventional boilers applies to the potential corrosion and erosion of inbed boiler tubes. It will be important to follow a schedule of cleaning and inspection to assure long boiler tube life.

2.4 SYSTEM PERFORMANCE - NO_x CONTROL

2.4.1 Factors Affecting NO_x Formation and Emission Reduction

NO_x emitted during AFBC coal combustion is virtually all in the form of NO. Argonne National Laboratory has found that NO accounts for 98 percent or more of the total NO_x emission.[75] In tests by Pope, Evans, and Robbins (PER), oxides of nitrogen other than NO were found to average between 10 to 30 ppm.[76] The high proportion of NO has also been verified in experimentation at MIT.[77]

Design and operating factors which influence the formation and control of NO_x in atmospheric fluidized-bed combustors include:

- Temperature
- Excess air
- Gas residence time
- Fuel nitrogen
- Factors affecting local reducing conditions
- Coal particle size

The kinetics of NO_x reduction are not well defined at this point and actual reductions cannot be predicted based on variation of different operating parameters. In some cases, different investigators report conflicting results relative to the influence of parametric variations.

2.4.2 Temperature

In the range of FBC operating temperatures (800° to 900°C), there is little correlation between temperature and NO_x emission. Westinghouse has compiled existing NO_x data as part of a comprehensive statistical study to determine the behavior of FBC with regard to NO_x and to develop a predictive mathematical model. A five-term nonlinear regression equation was developed based on equivalence ratio,* and temperature. Comparison of the model and actual data at an excess air rate of 18 percent is shown in Figure 12. A peak is seen between 800° and 900°C and emission rate falls off at temperatures below and above this range.

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*Actual fuel-to-air ratio ÷ stoichiometric fuel-to-air ratio.

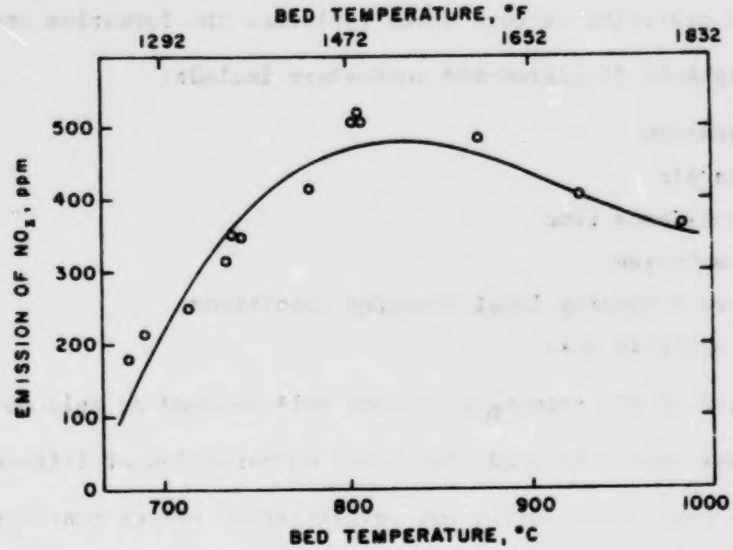


Figure 12. NO_x versus bed temperature, equivalence ratio 0.847 (18 percent excess air).

A temperature maximum for NO_x emissions was found by Pereira, Beer, and Gibbs at 750° to 800°C.[78] They concluded that NO_x emissions increased with temperature up to about 750°C because of a decrease in NO reduction by CO, hydrogen, and unburned hydrocarbons. At temperatures greater than 800°C, NO_x reduction by char is accelerated and emissions again decrease. Above 900° to 1000°C, thermal NO_x formation becomes significant and the emission rate of NO_x begins to increase.

PER has performed several tests at elevated temperature in their fluidized-bed module (FBM).[79] The results shown in Figure 13 are scattered but a definite upward trend exists. In the probable AFBC operating temperature range shown, the maximum NO_x emission rate is about 230 ng/J (0.53 lb/10^6 Btu) but the average is about 200 ng/J (0.47 lb/10^6 Btu).

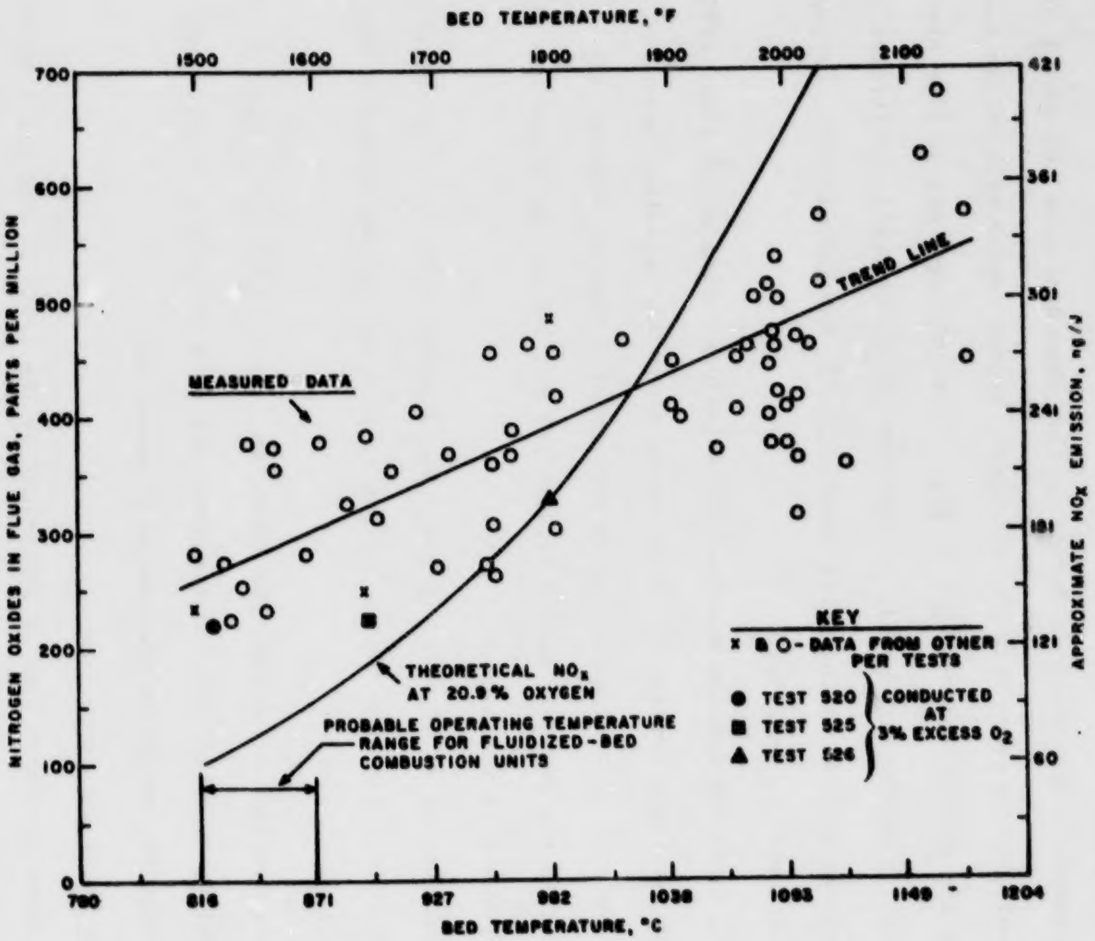


Figure 13. NO$_X$ emission rate as a function of bed temperature based on testing in the PER Fluidized-Bed Module (FBM).[79]

Several other investigators have reported similar results[80,81,82] with regard to increased NO_x emissions at elevated operating temperature.

2.4.3 Excess Air

The amount of NO_x which is formed is also dependent upon the amount of oxygen available to react with the nitrogen. Excess oxygen will attack nitrogen compounds and convert the nitrogen to NO_x. Thus, nitrogen may be liberated which might normally remain fuel-bound, in addition to NO_x liberated by thermal fixation. Several studies support this concept.[83,84] Substoichiometric oxygen results in lower NO_x emissions.

Pope, Evans and Robbins measured NO_x emissions during pilot plant (FBC) and full-scale (FBM) testing.[85] Pilot plant (FBC) NO_x emissions increased from 320 ppm at 1 percent oxygen up to 440 ppm at 5 percent oxygen. NO_x emissions from the FBM unit ranged between 280 and 340 ppm as oxygen was increased from 1 to 4 percent.

During testing at Argonne, in the 6 in. diameter unit, NO_x concentration was found to increase from 400 to 500 ppm as flue gas oxygen increased from 2.6 to 11.8 percent.[86]

During testing of the CBC, PER found NO_x levels to be independent of excess air. However, at normal FBC temperatures, the bulk of testing results support the fact that NO_x emissions increase with excess air.

2.4.4 Gas Phase Residence Time

Gas phase residence time is determined by the ratio of bed depth and fluidization velocity. For constant bed depth, gas residence time is inversely proportional to fluidization velocity. Jonke, et al., found an inverse relationship between NO_x emission reduction and fluidization velocity.[87] The results suggest that NO_x control is improved at longer gas phase residence times, probably because more time is available for the reduction of NO to elemental nitrogen.

2.4.5 Fuel Nitrogen

Testing which has been performed to date indicates that most of the NO emitted from AFBC evolves from conversion of fuel nitrogen. In fluidized-bed combustion, total NO emissions are greater than the equilibrium concentration expected based on thermal fixation of atmospheric nitrogen, represented by the following reaction:

$$N_2 \text{ (atmospheric)} + O_2 \rightarrow 2NO \qquad (1)$$

This additional NO is attributed to conversion of fuel nitrogen, or:

$$2N \text{ (fuel)} + O_2 \rightarrow 2NO \qquad (2)$$

Studies at ANL predicted thermal NO_x (reaction 1) formation of only 100 ppm,[88] however, measured emissions average about 350 ppm at normal FBC temperatures.[89] In other experimentation at Argonne, air nitrogen was replaced with argon, and no significant difference was found in NO_x emission rates. These experiments indicate the significance of reaction 2 (fixation of fuel nitrogen) in NOx formation in AFBC boilers. In atmospheric FBC, as much as 90 percent of the NO_x is formed from the nitrogenous compounds in the coal, and 10 percent is due to the fixation of nitrogen from the combustion air.[90]

2.4.6 Factors Affecting Local Reducing Conditions

Although most of the NO emitted is derived from fuel nitrogen, there is very little correlation between fuel nitrogen content and total NO_x emission rate, apparently because of other interactions in the bed. The most important point is that NO_x is formed near the bottom of the bed and is reduced to elemental nitrogen as it rises through the bed.[91] If all the nitrogen in a coal of 1.4 percent N content were converted to NO, 2,500 ppm would be emitted.[92] Since average NO emissions are generally much lower than this (300 to 600 ppm)[93] it appears that the chemical NO reduction mechanism overrides any variation that would result in NO_x emissions during fluidized-bed combustion of coals with varying nitrogen concentrations.

Evidence of this NO formation and reduction mechanism has been found in several studies. ESSO Research found that by adding 250 ppm NO to the combustion air, NO_x emissions only increased by a few ppm.[94] Pope, Evans, and Robbins noted a decrease in NO_x concentration between samples at increased heights above the fluid bed, also possibly indicating that a reduction reaction was taking place after the formation of NO_x. They report that the reduction of NO between the bed and the stack is as great as 45 percent. Dilution from air leakage accounted for only 15 percent of the reduction.[95] Results of studies at MIT also show a correlation between NO concentration and height above the air distributor plate.[96] Figure 14 illustrates NO concentrations measured at the wall and center line of a 30 × 30 cm combustor at two different operating temperatures.

A likely NO_x reduction mechanism in FBC is:

$$2CO + 2NO \rightarrow 2CO_2 + N_2 \qquad (3)$$

Carbon monoxide in the bed reduces NO to elemental nitrogen, with reduction dependent on gas phase residence time, temperature and other bed characteristics.[97] At higher temperatures, lower quantities of CO are available to reduce NO, so that final NO emissions are greater.

Another reduction reaction which may be taking place in FBC is a bit more complex. Investigators at Argonne National Laboratory observed that NO and SO_2 react over a partially sulfated lime bed, but that no reaction between the two occurs over pure $CaSO_4$ or pure CaO.[98] Figure 15 illustrates the relationship between sorbent feed rate and NO_x emission rate determined by investigators at ANL.

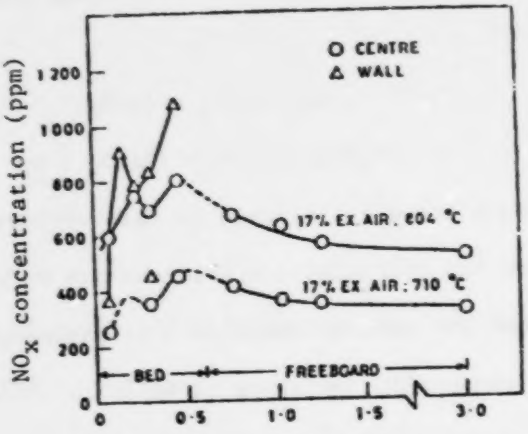


Figure 14. NO concentrations at different levels above distributor plate of 30 × 30 cm combustor reported by Massachusetts Institute of Technology.[96]

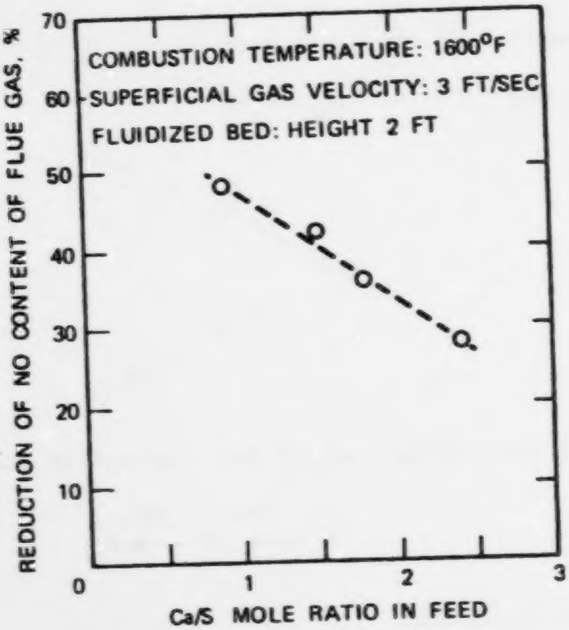


Figure 15. Reduction in NO versus Ca/S (ANL).[98]

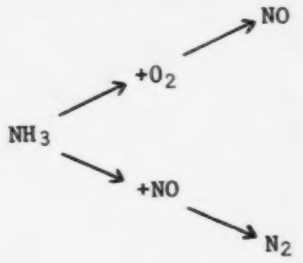
The reactions are assumed to be the following according to ESSO Research and Engineering:[99]

$$CaO + SO_2 \rightarrow CaSO_3 \qquad (4)$$

$$2CaSO_3 + 2\ NO \rightarrow 2CaSO_4 + N_2 \qquad (5)$$

These reactions were found to increase in rate with temperature decreases. Temperatures below FBC operating temperatures are more conducive to the reaction. This indicates that the rate of reaction 3 is probably greater than that of reaction 5 under normal FBC operating conditions. ESSO Research and Engineering reported that NO was reduced 20 to 40 percent over a partially sulfated bed as compared to an inert bed.[100] Battelle Columbus Laboratories also reported a 27 percent decrease in NO emissions over a partially sulfated bed versus an inert bed.[101]

NO may also be reduced to elemental nitrogen by reaction with coal volatiles, especially ammonia.[102] Fuel nitrogen, exemplified by ammonia in this case, takes part in two parallel reactions:



where NO is an intermediate in the two consecutive reactions:

$$O_2 \xrightarrow{NH_3} NO \xrightarrow{NH_3} N_2$$

2.4.7 Coal Particle Size

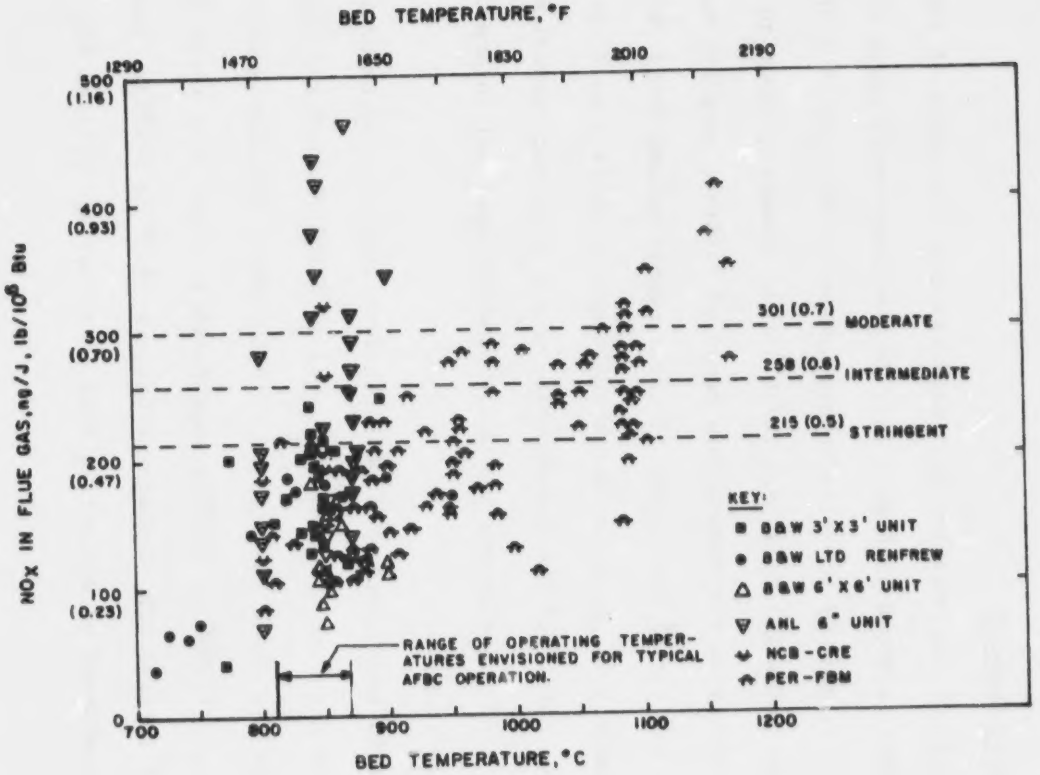
The effect of coal particle size on NO_x emissions is unclear. The National Coal Board compared NO_x emissions between systems using -3175 micron coal and -1680 micron coal. The results show NO_x reduction of 100 ppm as the coal size was reduced.[103] Investigators at Westinghouse have concluded, however, that smaller coal feed particles cause an increase in NO_x emissions.[104] Further testing is required to determine which conclusion is valid. Pereira and Beér[105] found that reduction of NO to elemental nitrogen significantly increased as char particle size decreased.

2.4.8 NO_x Emission Data Summary

A composite diagram of NO_x emission data measured over the range of normal FBC operating conditions is shown in Figure 16. In the temperature range of interest (800° to 900°C), most of the data points are below 260 ng/J (0.6 lb/10^6 Btu) and about half are below 215 ng/J (0.5 lb/10^6 Btu). However, about 10 percent of the test results in the temperature range of interest show NO_x emissions above 300 ng/J (0.7 lb/10^6 Btu). All of these higher values (>0.7 lb/10^6 Btu) are from the Argonne 6 in. diameter bench-scale unit. It is significant to note that all of the data from the larger units measured during operation at envisioned typical AFBC temperatures are well below the optional intermediate and stringent levels of control.

PER has made several measurements of NO_x emissions from the FBM experimental unit. Although much of the data are above 215 ng/J (0.5 lb/10^6 Btu) and about one-quarter of the data are above 301 ng/J (0.7 lb/10^6 Btu), the measurements were made at temperatures (1000° to 1200°C) significantly above that range expected in operation of typical AFBC industrial boilers (815° to 870°C). Therefore, they were not considered as supporting data in selecting optional NO_x control levels for AFBC.

These data are reported from experimentation where there was generally no intentional variation of design or operating conditions to reduce NO_x emission. This indicates that larger industrial AFBC boilers should be capable of meeting



*THESE POINTS ARE ESTIMATED FROM DATA REPORTED IN ppm, THUS THE ACCURACY OF THESE POINTS IS ASSUMED TO BE ± 30%

Figure 16. Summary of NO_x emission data from experimentation in AFBC test units.

NO_x levels as low as 215 ng/J (0.5 lb/10^6 Btu). If gas residence times are increased to enhance SO_2 control, this should aid in lowering NO_x emissions even further.

2.4.9 Potential Methods of Enhancing NO_x Control in AFBC Boilers

An alternative operating mode that can be used to reduce NO_x emissions further is two-stage combustion. This method can be applied to conventional boilers, and some preliminary testing has been conducted on FBC units. The combustion air is fed into the boiler in two stages. In the initial stage, near stoichiometric air is fed into the fluidized bed. Secondary air is fired into the boiler above the bed. In this stage, the burner must be carefully controlled in order to give minimal NO_x formation. In conventional combustion, two-stage combustion provides an effective reduction of about 30 to 50 percent thermal NO and up to 50 percent fuel derived NO_x.[106] Further testing is required in order to define the NO_x control potential of two-stage combustion in FBC systems.

Some of the most recent work at the SATR, EXXON, the Battelle MS-FBC, and the Enköping district heating plant are of interest because of the diversity in design, size, and results. The SATR is a small AFBC pilot plant, mainly designed for investigating particulate control. The EXXON miniplant is a small pressurized unit. The MS-FBC is a small recirculating bed FBC. The Enköping FBC is a two-stage combustor located in Sweden which generates 38,600 kg/hr (85,000 lb/hr) steam. Only the Enköping unit is designed as a staged combustor.

Staged combustion at the SATR reduced NO_x emissions to the 100 to 200 ppm range.[107] During initial trials SO_2 emissions increased somewhat. Subsequent testing with altered conditions reduced SO_2 emissions to below 200 ppm at a Ca/S ratio between 3.5 to 4, while maintaining low NO_x levels. No estimates of combustion efficiency are available.

The testing at EXXON resulted in substantial reductions of NO.[108] In one test, emissions of 0.05 lb/10^6 Btu were attained.* Unfortunately, both sulfur retention and combustion efficiency suffered. Sulfur emission reduction dropped from 74 percent down to 47 percent removal. Combustion efficiency dropped from 95 down to 90 percent.

Preliminary testing in the MS-FBC resulted in NO_x emissions dropping from 400 ppm to 150 ppm.[109] No change in sulfur capture or combustion efficiency was noted. One possible explanation for the good results obtained is the presence of an entrained bed throughout the freeboard. The freeboard is maintained at 1550°F to maximize sulfur capture. Thus, staged combustion helps maintain freeboard temperature for sulfur capture while reducing NO_x emissions.

No data on coal combustion in the Enköping unit are available, although results of a preliminary coal test in spring 1978 were made available to the U.S. EPA.[110] Sulfur capture of 75 percent at a Ca/S of 1.5, virtually 100 percent combustion, and very low NO_x are claimed for the unit while operating on high sulfur oil with 5 percent excess air. Staged combustion is employed to improve combustion efficiency at low excess air levels.

Studies at Argonne and ESSO showed that significant reduction of NO_x could be achieved in fluidized-bed combustion by the application of two-stage combustion. Argonne's test showed 70 to 100 ppm NO using two-stage combustion, where under similar single-stage conditions they measured 180 to 500 ppm NO.[111] ESSO's data show a reduction of NO from 620 ppm at 110 percent air in single-stage combustion to 200 ppm NO when the same amount of air was fed in stages (43 percent primary, 67 percent secondary).[112]

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*This is a pressurized FBC reactor and the chemical kinetics may be different. The trend of the data, however, supports the phenomena hypothesized for atmospheric systems.

Reduction of NO_x by staged combustion may be due to several reasons. In the primary stage there is insufficient oxygen to react with the nitrogen, and under substoichiometric reducing conditions, there is a greater amount of unburned fuel present, primarily in the form of CO, which reduces NO to N_2.

MIT has made several recommendations for combustion modifications for NO_x control based on small laboratory fixed and fluid bed experimentation.[113] Among the optional operating techniques postulated are:

- Inject 10 percent or more of the stoichiometric combustion air as secondary air into the freeboard for NO reduction by char in the bed and complete combustion of CO in the freeboard.
- Inject recycled char close to the top of the bed to promote the decomposition of NO rising through the bed.
- Inject recycled char together with coal and sorbent into a shallow uncooled bed situated above the main bed (see Figure 17) to reduce NO and produce favorable conditions for volatile combustion and sulfur retention in the "top fed" fluidized combustor.

The performance and economics of such options must be further assessed.

Further investigation of two-stage combustion in large scale FBC units is required to ensure that suitable SO_2 control and combustion efficiency can be attained simultaneously with low NO_x emissions. Another item to investigate is tube corrosion brought on by possibly shifting oxidizing/reducing zones in the unit.

Other techniques which could be considered for further NO control in AFBC include flue gas recirculation and ammonia/urea injection. Further testing is required to determine the incremental NO reduction which can be expected under these optional operating conditions.

If further reduction of NO_x is necessary, catalytic reduction is a possible approach. Studies have been done using various metal oxides and metal powders.

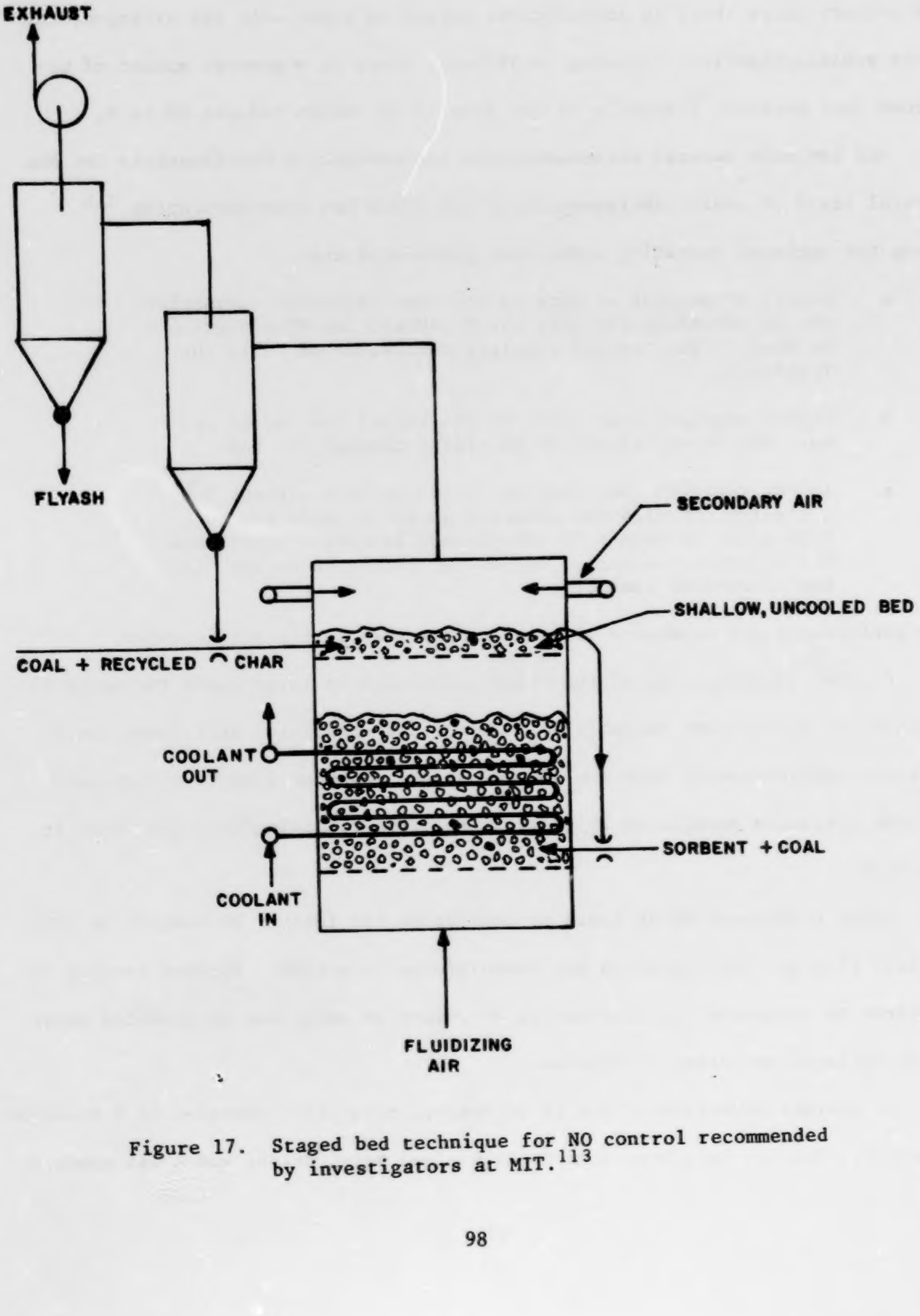
EXHAUST
FLYASH
SECONDARY AIR
SHALLOW, UNCOOLED BED
COAL + RECYCLED CHAR
COOLANT OUT
SORBENT + COAL
COOLANT IN
FLUIDIZING AIR

Figure 17. Staged bed technique for NO control recommended by investigators at MIT.[113]

Al_2O_3 and ZrO_2 seemed to have no effect on NO_x formation, and cobalt oxide seemed to increase NO_x.[114] However, the addition of nickel powder to the reaction chamber showed a significant decrease in NO_x formation.[115] This particular catalyst is extremely expensive and economically unrealistic for use in FBC, yet the study does demonstrate the feasibility of using a catalyst for NO_x control.

2.5 SYSTEM PERFORMANCE — PARTICULATE CONTROL

2.5.1 FBC Boiler Design Parameters Affecting Particulate Emissions

The most important design factors influencing the quantity of particulate emissions from an atmospheric FBC can be grouped as follows:

- Coal
  - ash content
  - sulfur content
  - agglomeration characteristics
- Sorbent
  - particle size
  - attrition and decrepitation characteristics
- Operation
  - superficial velocity
  - primary recycle
  - use of carbon burnup cell
  - additives
- Bed Geometry
  - cross sectional area
  - bed depth
  - orientation of boiler tubes
  - grid design
  - freeboard

2.5.1.1 Coal Type--

The type of coal used in an FBC boiler will influence the quantity and size distribution of stack gas particulate emissions. The most important factors are coal ash content, coal sulfur content and ash agglomeration characteristics. Fly ash emissions will increase with increasing ash content since it is reported that virtually 100 percent of all coal ash is elutriated from the fluidized bed.[116] Particulate emission data analyzed by Babcock & Wilcox showed very little correlation between emission rate and coal or additive particle size.[117]

Particulate emissions will increase with increasing sulfur content because of greater sorbent requirements for SO_2 control. Although most of the spent sorbent is likely to be withdrawn from the bed, increasing sorbent feed rate may be expected to increase the amount of sorbent elutriated.

Collection of elutriated ash by primary recycle cyclones will be influenced by ash agglomeration. The temperature in the fluidized bed is lower than that associated with ash agglomeration in conventional systems, but if this does occur in a fluidized system, the internal cyclones will provide highly efficient capture of large-sized agglomerated material elutriated from the bed.

2.5.1.2 Sorbent Type--

In fluidized-bed combustion, sorbent material can represent a significant portion of the particulate material reaching the final control device. The amount of sorbent elutriated depends upon sorbent size distribution and the relationship between the terminal particle settling velocity and superficial fluidization velocity. Any change in sorbent particle size which results in terminal particle settling velocities less than superficial velocity will tend to cause elutriation of that size fraction. There is also a possibility of emitting particles with higher terminal velocities due to the complex nature of a fluidized system,[118] however, higher freeboard designs will help reduce carryover of "splashed" coarse particles. In addition to immediate sorbent fines elutriation upon sorbent feeding, two mechanisms are responsible for *in situ* reduction of sorbent particle size, including:

- Decrepitation
- Attrition

Fines are formed as a result of sorbent decrepitation during calcination and/or sulfation. Sorbent particles are roasted and cracked into finer size fractions, the extent of which depends on sorbent type.

Attrition refers to mechanical grinding of sorbent particles as a result of turbulent particle interactions in the bed. This phenomenon occurs most rapidly during calcination and can cause a significant increase in total particulate emissions if proper sorbents are not used.

2.5.1.3 Operating Conditions--

The role of superficial velocity in particulate elutriation is pointed out above in the discussion of sorbent characteristics. In general, particulate emissions from the FBC will increase directly with increasing superficial gas velocity.

The use of primary recycle to enhance combustion efficiency and SO_2 control efficiency (by allowing for longer carbon and sorbent residence times) provides significant reduction of particle loading to the final particulate control device.

Another significant operating factor affecting particulate emissions from an FBC system is use of a separate carbon burnup cell (CBC) to burn recycled carbon elutriated from the main combustor. The CBC differs from the FBC in many respects, including the following:[119]

- Characteristics of combustion material (i.e., finer than FBC feed, higher proportion of ash, lower proportion of carbon
- Higher temperature operation, ~1100°C (2000°F)
- Higher excess air, ~50 percent
- Lower fluidizing velocity

Particulate emissions from the CBC will decrease with increasing temperature, as discussed in Section 2.5.3.2. Although this is the case, we do not expect widespread use of carbon burnup cells in industrial boilers.

Salt additives can be used to increase sulfur retention in the bed. Studies by PER indicated that particulate emissions increased during salt addition.[120] Their investigation also noted that attrition loss was more severe during startup and salt addition.[121]

2.5.1.4 Bed Geometry--

The quality of fluidization is directly related to bed depth, with gas bypassing, slugging, and bubbling decreasing as bed depth increases. As a result, particle elutriation is also minimized at increased bed depths.[122]

Bed diameter and boiler tube configuration also influence fluidization characteristics. The quality of fluidization increases with increasing bed diameter, and indicates that full-scale units will have better fluidization characteristics than bench- and pilot-scale units currently in operation.[123]

Boiler tubes in the bed can serve to break up gas bubbles and provide smoother fluidization. Tubes should be oriented to allow for good mixing. Definitive guidelines for boiler tube orientation have not been developed, but many operating pilot plant units incorporate horizontally-mounted tubes. Planned units are considering inclined tubes to allow for natural coolant circulation. Tube packing also has an effect, causing large temperature gradients if packed too closely. This is a sign of poor mixing.

Boiler tubes also act as baffles, both water tubes submerged below the surface of the bed, and convective tubes in the freeboard above the bed. Such a baffling effect could reduce the amount of particles elutriated.

Grid design is another important factor in assuring proper mixing and fluidization. Uneven gas distribution may cause channeling and possible deactivation in portions of the bed. Designing grid pressure drop at approximately 40 percent of total bed pressure drop should provide for uniform gas distribution and mixing.[124] This will minimize particle elutriation due to gas bypassing and slugging or bubbling. Air distributor grid jets also contribute to attrition and emission of sorbent particles.

It is expected that future FBC designs will incorporate deeper beds, and smaller sorbent particles to improve SO_2 control (see Section 3.0). If so, expansion of the freeboard dimension will be important to avoid excessive particle elutriation. Some designers, most notably Babcock and Wilcox, have already worked higher freeboards into their designs.

2.5.2 FBC Boiler Operating Factors Affecting Particulate Control Device Performance

Selection and performance of a final particulate control device will depend on flue gas characteristics and particle characteristics as determined by basic boiler operating parameters. Control devices which could be used include ESPs, fabric filters, scrubbers, and cyclones. The use of each of these techiques is discussed below to the extent that the application differs from a conventional boiler/particulate control system.

2.5.2.1 Electrostatic Precipitators--

A hot-side or cold-side electrostatic precipitator (ESP) could be used as a final particulate control device in an FBC system. These options are illustrated in Figure 18.[125] The decision is based largely on particle resistivity, which is influenced by:

- Flue gas temperature
- Particulate carbon and alkali concentration, and SO_3 concentration in the flue gas
- Use of separate carbon burnup cell
- Use of additive
- Trace element concentration

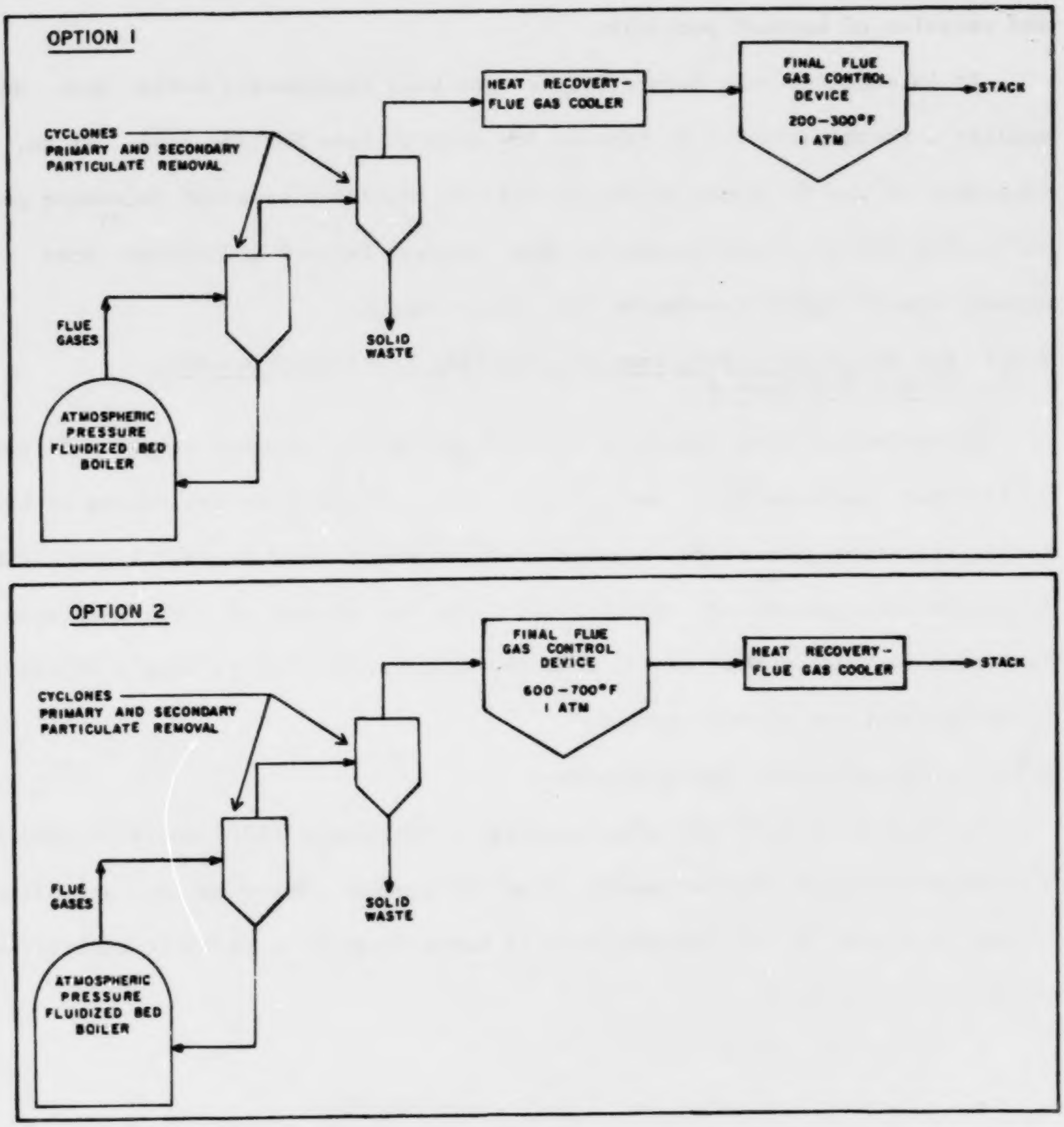
OPTION 1
HEAT RECOVERY-FLUE GAS COOLER
FINAL FLUE GAS CONTROL DEVICE
200–300°F
1 ATM
STACK
CYCLONES
PRIMARY AND SECONDARY PARTICULATE REMOVAL
FLUE GASES
SOLID WASTE
ATMOSPHERIC PRESSURE FLUIDIZED BED BOILER
OPTION 2
FINAL FLUE GAS CONTROL DEVICE
600–700°F
1 ATM
HEAT RECOVERY-FLUE GAS COOLER
STACK
CYCLONES
PRIMARY AND SECONDARY PARTICULATE REMOVAL
FLUE GASES
SOLID WASTE
ATMOSPHERIC PRESSURE FLUIDIZED BED BOILER

Figure 18. Control of particulate emissions from an atmospheric pressure FBC boiler.[125]

The resistance of particulate material should be in the range of 1×10^7 to 2×10^{10} ohm-cm for high efficiency performance.[126]

Analysis of particulates emitted from fluidized-bed combustion systems indicates that resistivity may be above the range required for acceptable ESP performance, especially at temperatures of 95° to 150°C (200° to 300°F) characteristic of cold-side control operation. Figure 19 illustrates a compilation of resistivity measurements made by TVA and Pope, Evans, and Robbins.[127] None of the data indicate that a cold-side ESP would function well, unless the TVA in situ measurements with limestone are extrapolated to temperatures of 120°C (250°F) or less, which is below the normal cold-side temperature range. Five data points at 315°C (600°F) are below 1×10^{10} ohm-cm indicating possible hot-side ESP control. Extrapolation to higher temperatures between 315° to 370°C (600° to 700°F) shows potentially lower resistivities.

There are several reasons why particle resistivity is a problem in fluidized-bed combustion. Very low concentrations of SO_3 have been recorded in FBC flue gas, and SO_3 appears to be of major importance in lowering the resistivity of fly ash collected by cold-side precipitators. All sorbent materials ($CaCO_3$, CaO, MgO, $CaSO_4$) have high resistivities. (Carbon content of the fly ash, on the other hand, could tend to lower resistivity.) Trace element distribution on fly ash particles from FBC could alter the volume conduction effect, an important factor in hot-side ESP operation.[128] The test data shown in Figure 19 are for emissions from the primary combustor with combustion efficiency in the range of 85 to 90 percent. In actual operation, combustion efficiencies as high as 95 to 97 percent may be approached, so that carbon concentrations in the flue gas will be reduced in comparison to this data.

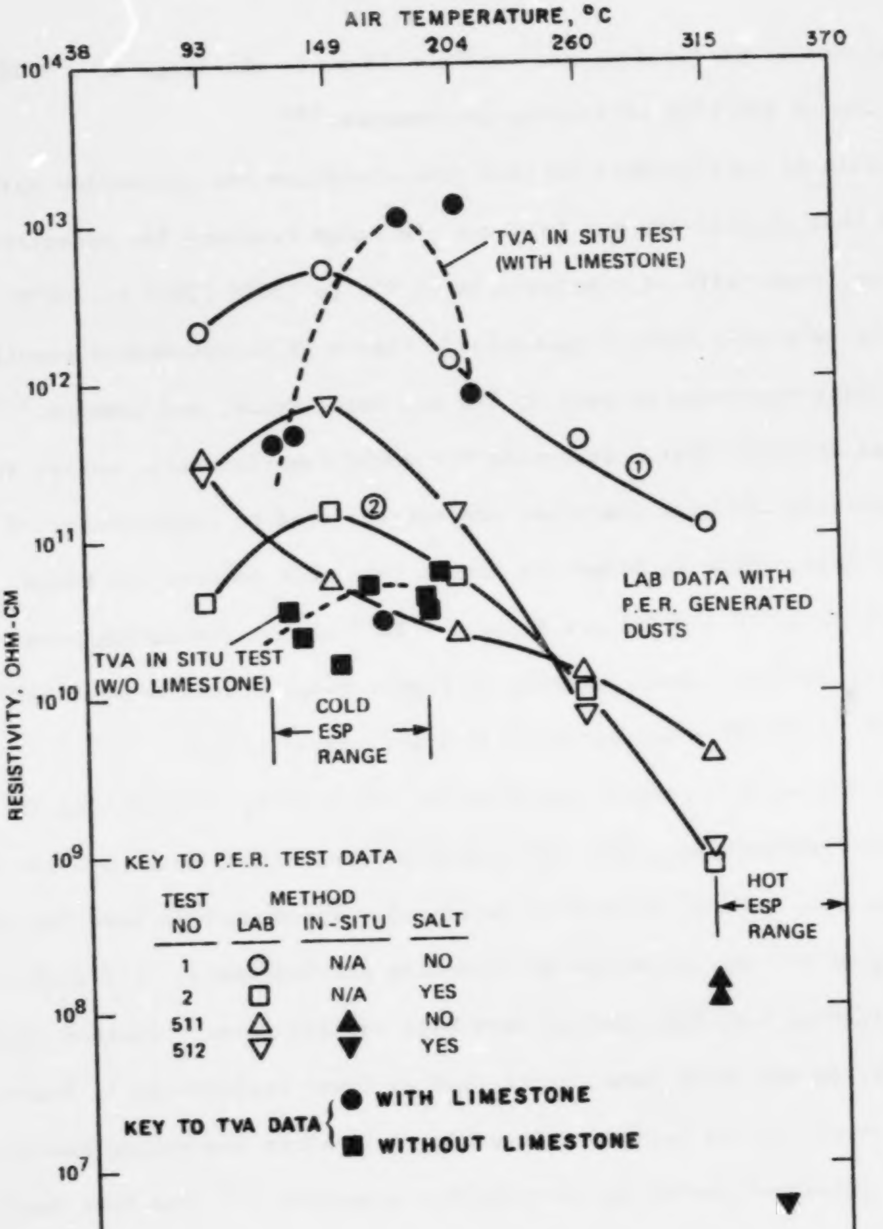

Figure 19. Resistivity of fluidized-bed particulate emissions.[127]

Carbon has high conductivity and, therefore, reduces resistivity. Thus, in full-scale industrial units, actual resistivities may be higher than reported in this testing.

2.5.2.2 Fabric Filters--

It is anticipated that fabric filtration technology will be readily adaptable and successful in controlling particulate emissions from coal-fired FBC boilers. Depending on the gas moisture (which should be low) slight problems could develop with pH of material captured in the filter, or lime hydration could cause temperature excursions or blinding at the fabric surface. The potential for bag-fires must also be considered due to uncertainty regarding the extent of carry-over of unburned carbon.

Water vapor in flue gases from combustion is primarily a result of the fuel hydrogen content and it produces a dew point of 50° to 60°C (122° to 140°F) at normal excess air. However, the SO_3 concentration (usually 1 to 2 percent of the SO_2 concentration) in a conventional coal-fired boiler raises the flue gas dew point. Equipment designed to collect dry particulate (fabric filters and dry electrostatic precipitators) must operate above the acid dew point. Most conventional coal-fired plants maintain flue gas temperatures between 150°C (300°F) and 180°C (356°F) to avoid corrosion problems. Robinson, et al.,[129] found that the Pope, Evans, and Robbins - fluidized-bed pilot plant produced an SO_3 concentration of 39 ppm when sorbent was not used, and no measurable SO_3 when sorbent was used. (Note: These early SO_3 results represent limited data and must be confirmed by further SO_3 analyses on other fluidized-bed combustors.) This low SO_3 concentration in the presence of sorbent, if confirmed, means that flue gases might be cooled to 95°C (200°F) or below for dry particulate collection and increased heat recovery.[130] The major problem in using fabric filters on conventional coal-fired boilers has been SO_3 and H_2SO_4 induced deterioration of the fabric. Therefore, fabric filter technology may be readily applicable to fluidized-bed combustion systems if the low SO_3 concentrations are confirmed.

2.5.2.3 Wet Scrubbers--

Wet scrubbers for final particle control application in FBC have not been seriously considered in this report because of the wet sludge/wastewater handling and disposal problem which would result. Since other particle control systems are anticipated to perform adequately on FBC and because an inherent attraction of FBC is dry waste production, wet scrubber use would probably not be considered by the industrial customer. In the event that scrubbers were used, they would have to be operated at high pressure drop with attendantly high power consumption and operating cost to provide high efficiency removal of fine particles.

2.5.2.4 Multitube Cyclones--

Multitube cyclones, which represented the most common type of inertial collector used for fly ash collection before stricter emission regulations were enacted, depend upon centrifugal forces (i.e., inertial impaction) for particle removal. They consist of a number of small-diameter cyclones (~5 to 30.5 cm diameter) (~2 to 12 in. diameter) operating in parallel and having a common gas inlet and outlet.

Fly ash collection by multitube cyclones is a well-established technology that has been applied for many years on all types of conventional coal-fired industrial and utility boilers. However, because of efficiency limitations they are now used mainly as precleaning devices.

A cyclone or multiple cyclones would be required to operate at high velocity to provide significant removal of fine particles. Table 15 shows typical efficiencies of three different cyclone collectors.[131]

TABLE 15. DISTRIBUTION BY PARTICLE SIZE OF AVERAGE COLLECTION EFFICIENCIES FOR VARIOUS PARTICULATE CONTROL EQUIPMENT[131]

Type of collector	Collection efficiency, %				
	Particle size range, µm				
	<5	5 to 10	10 to 20	20 to 44	>44
Simple cyclone	7.5	22	43	80	90
Multitube cyclone (12 in. diameter)	25	54	74	95	98
Multitube cyclone (6 in. diameter)	63	95	98	99.5	100

Removal of fines <5 to 10 µm probably would not be adequate with use of any of these cyclone arrangements, and if so, only at very high cost. Figure 20 illustrates comparative collection efficiencies for two axial-entry cyclones applied to conventional boilers with diameters of 15.2 and 30.5 cm (6 to 12 in.), respectively, as a function of percent of dust under 10 µm.

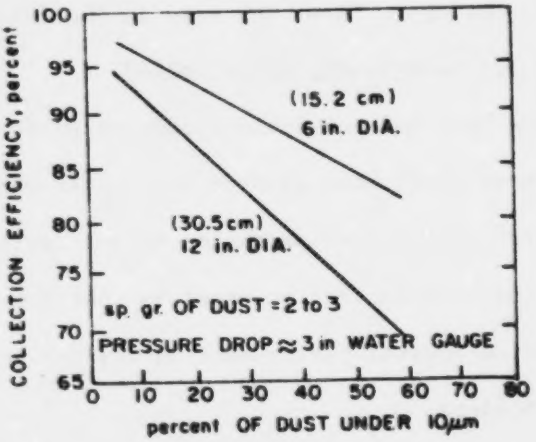

Figure 20. Typical overall collection efficiency of axial-entry cyclones.

The average inlet particle size to the final device in FBC is expected to range between 5 to 20 μm. If it is actually 10 μm or below, the maximum efficiency which could be expected based on this data (for conventional firing) is 73 and 85 percent for 30.5 cm (12 in.) and 15.2 cm (6 in.) diameter cyclones, respectively.

Although a great deal more testing is required in large-scale FBC systems to assess multitube cyclone performance capability, it is apparent that multitube cyclones would probably only be adequate for moderate particulate control levels.

2.5.3 Particulate Emission Data from AFBC Units

Actual test data demonstrating the efficiency of final particulate control devices applied to coal-fired atmospheric FBC boilers are not available. Particulate emission data which do exist generally represent loadings in the flue gas to, and the exhaust from, primary cyclones applied to the FBC or CBC. To date, large FBC units have not operated long enough to demonstrate final particulate control technology. Thus, the data on the following pages represent data from units which are essentially uncontrolled.

The factors affecting final particulate control performance, as they differ from conventional systems, have been pointed out. Although certain problems require further research, and actual particle control device performance on FBC must be demonstrated, the current prospect is that hot-side ESP or fabric filter use should provide control performance equivalent to applications on conventionally-fired boilers.

To support the probable adequate performance of final particulate control devices, available emissions data pertaining to exhaust from the primary cyclones is discussed below.

2.5.3.1 Particle Size Data--

Figure 21 illustrates particle size distributions measured for emissions from conventional and FBC boilers. The FBC particle size distribution was measured by PER in their 1.5 ft × 6 ft fluidized-bed module.[132] Isokinetic sampling was used along with an MSA particle analyzer for subsieve size particles and the data represent emissions at the inlet to the final control device. A 12-element multicone dust collector was used for primary particulate removal. In addition, large particle fallout occurred in the air preheater. Exact operating conditions during this FBM run are not known, but at this time the FBM was being operated at relatively high superficial velocities (3 to 4 m/sec). Bed depth was variable with a maximum slumped depth of about 0.25 m (30 in.). The distribution reported by Midwest Research Institute (MRI) represents emissions after a cyclone or similar mechanical collection device applied to a conventional pulverized coal boiler.[133] Particle sizing by MRI was performed using a Bahco classifier. Although this is limited data comparing a full-scale conventional system with a small FBC test system, it can be tentatively concluded that the size distributions of particulate emissions passing to final control devices in conventional and FBC systems are not radically different. It is possible, however, that particulate emissions from FBC may include a slightly higher concentration of fines.

Argonne has determined the particle size distribution of fines (by Coulter counter analysis) collected by their control equipment during two atmospheric FBC bench scale experiments in a 6 in. combustor. The operating conditions were as follows:[134]

- Temperature : 871°C (1600°F)
- Coal : -14 mesh Illinois, 4 percent S

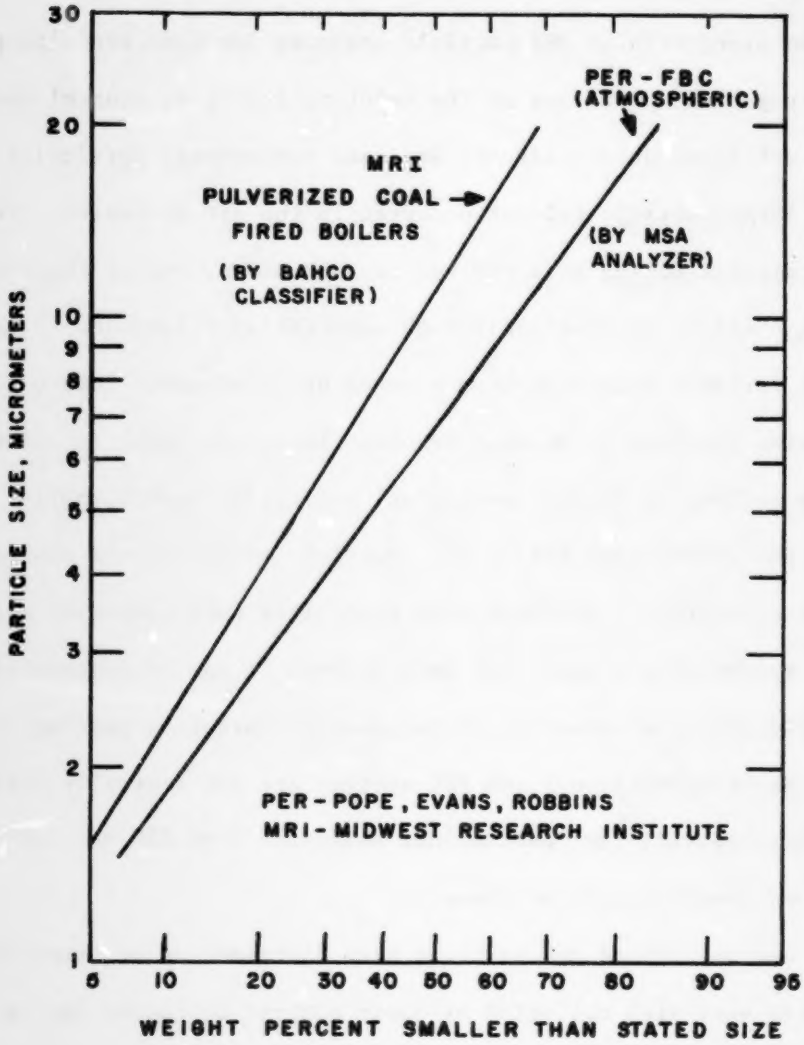

Figure 21. Particle size distribution before final control device.[132,133]

- Additive : BCR - 1359 calcined limestone
- Starting bed : 30 mesh alumina
- Bed height : static - 0.4 m (15 in.)
  fluidized - 0.6 m (24 in.)
- Superficial velocity: 0.9 m/sec (3 ft/sec)
- No recycle
- Ca/S : 2.4 to 2.9
- Excess air : 10 to 30 percent

The distributions are shown in Figure 22, and illustrate that the average size of particles collected in the primary cyclone is 70 μm. Particles exiting the secondary cyclone have an average size of 15 μm with about 3 percent <2 μm. Total collection efficiencies for the two devices were reported as 86 to 90 percent and 97 to 99 percent, respectively.

2.5.3.2 Emission Data

Table 16 presents a summary of particulate emissions data from PER, ANL, NCB, and B&W. PER conducted particulate emission testing during operation of the FBC and FBM test units in 1970.[135] (The FBC was a pilot-scale unit with a rectangular bed of 30 cm × 41 cm (1 ft × 1.3 ft) and the FBM was envisioned as a "full-scale" module with a rectangular bed of 46 cm × 183 cm (1.5 ft × 6 ft).)[136]

Testing downstream of the FBC cyclone indicated that about 10 percent of the fly ash escaped uncaptured. A summary of the results is shown in Table 16. During this test, a sintered ash bed 25 cm (10 in.) deep was operated at 843°C (1550°F) with 3 percent oxygen in the flue gas. Superficial velocity was not reported for this specific testing but it is known to have been varied between 1.8 to 4.3 m/sec (6 to 14 ft/sec) for all testing during this period. Fine sorbent (-325 mesh) was injected. PER concluded that the bulk of the sorbent

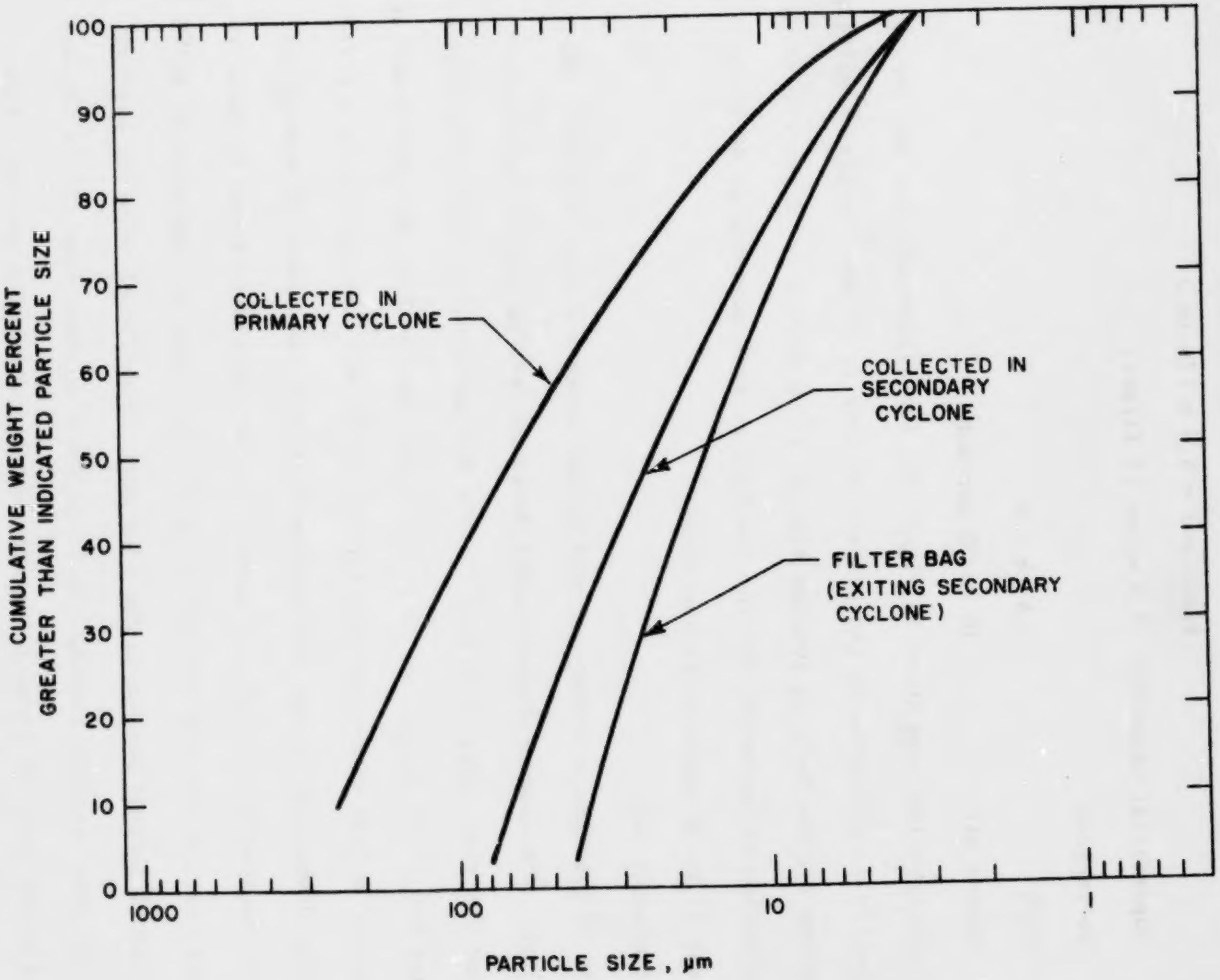

Figure 22. Typical particle size distribution of elutriated material measured at Argonne National Laboratory.[134]

TABLE 16. SUMMARY OF PARTICULATE EMISSION DATA, PRIMARY AND SECONDARY COLLECTION – ATMOSPHERIC FBC UNITS

Investigator	Coal parameters	Sorbent parameters	Other operating conditions	Particulate loadings	Primary and secondary collection efficiency
ANL[140] 6 in. bench scale FBC	4% sulfur Illinois coal. Coal ash feed rate -0.23 kg/hr (0.5 lb/hr)	BCR 1359 calcined limestone	Gas velocity of 0.9 m/s (3 ft/s); no recycle	Downstream of secondary cyclone: 0.45 g/m (0.198 gr/cf) or approximately:* 215 ng/J (0.5 lb/10^6 BTU)	86-90% primary cyclone 97-99% combined cyclones
ANL[141] 6 in. bench scale FBC	Coal 1.8-3.3 kg/hr (4.0-7.3 lb/hr)	Limestone input 0.5-1.1 kg/hr (1.1-2.3 lb/hr)		At combustor exit: 0.37-4.07 g/m (0.16-1.78 gr/cf) downstream of secondary cyclone: average-0.14 g/m (0.06 gr/cf) maximum-0.5 g/m (0.22 gr/cf)	Approximately 90% for combined primary and secondary cyclones
NCB[142] BCURA and CRE pilot scale combustors	Several varieties of low and high sulfur coal: 9-227 kg/hr (20-500 lb/hr)		Fluidizing velocity 0.61-3.35 m/sec (2-11 ft/sec)	Downstream of secondary cyclone: 0.23-1.37 g/m (0.1-0.6 gr/cf) or approximately:* 108-645 ng/J (0.25-1.5 lb/10^6 BTU)	95-98% for combined primary and secondary cyclones
PER[137] FBC 30 × 41 cm (12 × 16 in.)	Coal ash input kg/hr (lb/hr)	Limestone (1359) input kg/hr (lb/hr)	Sintered ash bed, 25 cm (10 in.) depth. 3% O_2 in flue gas Temp. = 843°C (1550°F)	Fly ash captured primary cyclone kg/hr (lb/hr) — Fly ash emitted kg/hr (lb/hr)	Approximately 90%
	5.8 (12.8)	0 (0)		10.0 (22.0) — 0.7 (1.5)	
	5.9 (12.9)	0 (0)		10.5 (23.2) — 1.1 (2.4)	
	5.7 (12.6)	9.7 (21.4)		18.6 (41.0) — 1.8 (3.9)	
	5.9 (12.9)	12.7 (28.0)		19.7 (43.4) — 2.2 (4.9)	
PER[137] FBM 0.46 × 1.8 m (18 × 72 in.)	Coal feed rate kg/hr (lb/hr)	Limestone feed rate kg/hr (lb/hr)	Limestone type (all sized at -44 μm)	Particulate emission rate after primary cyclone: Low§ ng/J* (lb/10^6 BTU)* — High§ ng/J* (lb/10^6 BTU)*	90-95%
	364-373 (800-820)†	164 (360)	Dolomite 1337 raw	318 (0.74) — 696 (1.62)	
	348 (765)†	100 (220)	Limestone 1359 raw	456 (1.06) — 718 (1.67)	
	400-420 (880-925)‡	60 (132)	Limestone 1337 hydrate	396 (0.92) — 494 (1.15)	
	345-364 (760-800)‡	128 (282)	Limestone 1337 raw	383 (0.89) — 602 (1.4)	
	327-339 (720-745)‡	44 (97)	Limestone 1359 raw	374 (0.87) — 598 (1.39)	
	336-345 (740-760)‡	30 (66)	Limestone 1359 hydrate	327 (0.76) — 473 (1.1)	

(continued)

TABLE 16 (continued).

Investigator	Coal parameters Coal input kg/hr	 (lb/hr)	Sorbent parameters Limestone input kg/hr	 (lb/hr)	Other operating conditions Type and limestone size (μm)	Particulate loadings Particulate at WS inlet ng/J*	 (lb/10^6 BTU)*	Primary and secondary collection efficiency
Babcock and Wilcox[145] 0.91 × 0.91 m (3 × 3 ft)	112-758	(248-758)	23-77	(52-170)	6350 × 0 (Lowellville)	2253-3375	(5.24-7.85)	Primary collection efficiency raged between 50-80%. This is low in comparison to efficiencies achieved when cyclones are used for primary fly ash removal.
	200-218	(440-480)	21-68	(47-150)	2380 × 0 (Lowellville)	2878-3689	(6.74-8.58)	
	209-222	(460-490)	20-66	(45-145)	1000 × 0	3078-4170	(7.16-9.70)	
	154-245	(340-540)	45-64	(100-140)	Pulverized (Lowellville)	5434-7825	(12.64-18.20)	
	134-240	(295-530)	19-49	(41-107)	44 × 0 ($CaOH_2$)	4970-7145	(11.56-16.62)	
	220-227	(485-500)	116-127	(256-279)	2380, 1000, pulverized Greer	3637-10,623	(8.46-24.71)	
	222-230	(490-507)	104-113	(230-250)	2380, 1000, pulverized Grove	3457-15,215	(8.04-35.39)	

Investigator	Test series	Coal input kg/hr	 lb/hr	Limestone input kg/hr	 lg/hr	Particle loading primary cyclone outlet ng/J	 (lb/10 BTU)	Primary collection efficiency
Babcock and Wilcox[144] 1.8 × 1.8 m (6 × 6 ft)	3-2	892	(1965)	261-277	(575-610)	3224	(7.5)	83
	4-1	818-890	(1801-1961)	245-291	(540-640)	3323-6453	(7.73-12.01)	50-65
	4-2	805-809	(1773-1783)	281	(620)	3130-3147	(7.28-7.32)	70
	4-3	847-903	(1866-1990)	300	(660)	3203-3431	(7.45-7.98)	65
	5-1	800-810	(1762-1785)	341	(750)	2042-2068	(4.75-4.81)	80
	5-2	823-894	(1813-1970)	295-409	(650-900)	770-1367	(1.79-3.18)	87-91
	5-3	923-956	(2033-2105)	281-302	(620-665)	2128-2205	(4.98-5.13)	76
	6-1	799-895	(1759-1971)	217-423	(478-931)	1638-2184	(3.81-5.08)	75
	6-2	727-898	(1601-1977)	160-267	(353-589)	1961-3276	(4.56-7.62)	61-72
	6-3	883-914	(1944-2014)	198-215	(437-473)	3603-3770	(8.38-8.77)	60

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*Estimated by GCA

†Ohio No. 8, unwashed coal - 4.5% S, 10.7% ash.

‡Ohio No. 8, washed coal - 2.6% S, 7.2% ash.

§High value measured during fine sorbent addition; low value measured with no sorbent addition.

NOTE: Limestone Type and Size = Lowellville limestone led at top size of 9510 μm for all testing.

was retained in the collector despite the -325 mesh particle size. About 10 percent of the input energy was lost as carbon in the fly ash. No attempt was made to recover this loss by fly ash recirculation.

Particulate testing was also conducted during several runs of the PER FBM unit.[137] Feed coal was Ohio No. 8. Sulfur concentration was 4.5 percent for unwashed coal and 2.6 percent for washed coal. The ash concentrations were 10.7 percent and 7.2 percent, respectively. Superficial velocity was approximately 3.4 m/sec (11 ft/sec) and sorbent feed particle size was -44 μm. Particulate emission measurements downstream of the primary cyclone are summarized in Table 6. PER reports that 52 percent by weight (90 percent by number) of particles exiting the cyclone were smaller than 5 μm. In all cases, cyclone collection efficiency exceeded 90 percent.

Use of carbon burnup cell in industrial FBC systems is not anticipated, however, measurements made by PER on their modified fluidized-bed column indicate the effect of operating temperature on particulate emissions. As shown in Figure 23,[138] particulate emissions decrease with increasing temperature, probably due to improved carbon combustion and ash agglomeration.[139] Over the temperature range tested, particulate emissions varied from 430 up to 3,440 ng/J (1 to 8 lb/10^6 Btu).

During the ANL studies of particle size distribution, a grain loading of 0.198 gr/cf (approximately 215 ng/J) was measured in the exhaust from the secondary cyclone.[140] ANL ran tests to determine cyclone efficiency (primary cyclone, 6-5/8 in. diameter; secondary cyclone, 4-1/2 in. diameter).[141] Flue gas volumes ranged from 3.8 to 6.6 lps (8 to 14 cfm), coal feed from 1.8 to 3.3 kg/hr (1.1 to 2.3 lb/hr). The dust loading in the combustor exhaust prior to both cyclones ranged from 0.16 to 1.78 gr/cf, (approximately 170 to 1,920 ng/J)

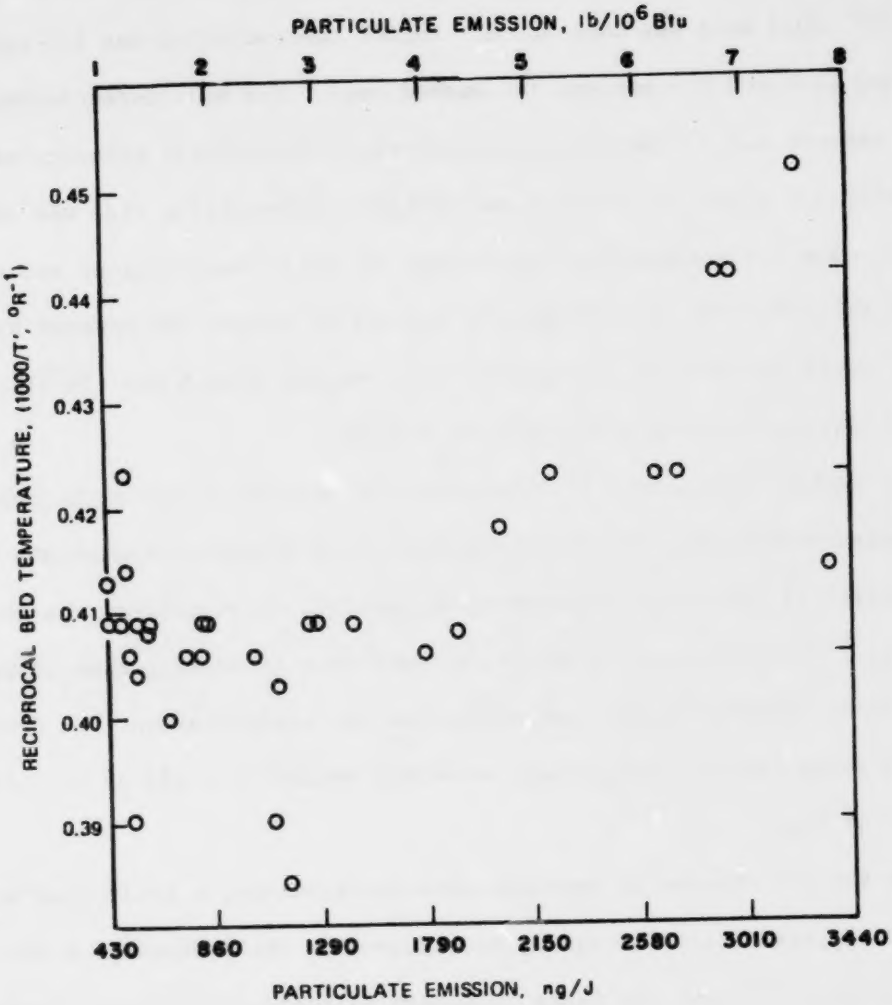

Figure 23. Particulate emissions as a function of temperature as determined by PER in simulated CBC operation.

and the average measured loading after the secondary cyclone was 0.06 gr/cf (approximately 65 ng/J) ranging to a maximum of 0.22 gr/cf (approximately 240 ng/J). Combined overall cyclone efficiency was approximately 90 percent.

Data obtained by the National Coal Board[142] (from the 1.5 ft × 3 ft CRE reactor) using primary and secondary cyclones with collection efficiencies of 90 percent for 10 μm particles, showed exhaust particulate loadings between 0.1 and 0.6 gr/scf (approximately 110 to 650 ng/J). This indicated a combined collection efficiency for the two cyclones of 95 to 98 percent. Fractional efficiencies for the two cyclones are shown in Figure 24, and show how drastically cyclone efficiency drops for particles smaller than 10 μm. During testing, superficial gas velocity ranged between 1.2 to 2.4 m/sec (4 to 8 ft/sec). The primary cyclone had a 0.6 m (24 in.) diameter with a height of 2.7 m (8 ft, 10 in.). The secondary cyclone had a diameter of 0.43 m (17.25 in.) and height of 2 m (6 ft, 7 in.).

The primary fines were sampled using an incremental sampler designed to take a full cross sectional sample of the entire fines flow. Samples of exhaust dust were obtained from a probe 1.2 m (4 ft) after the secondary cyclone by extracting isokinetically a known volume of exhaust gas and passing it through a weighed filter.

Babcock and Wilcox has compiled particulate emission data reported by several investigators and has found that the best correlation of particulate emission rate is based on superficial air velocity. Figure 25 illustrates the relationship between uncontrolled particulate emission rate and superficial air velocity, as reported for one specific sorbent type.[143] This particular graph is based on data from NCB (the 1.5 ft × 3 ft, 27 in. diameter, and the 6 in. diameter units) and ANL (the 6 in. diameter unit).

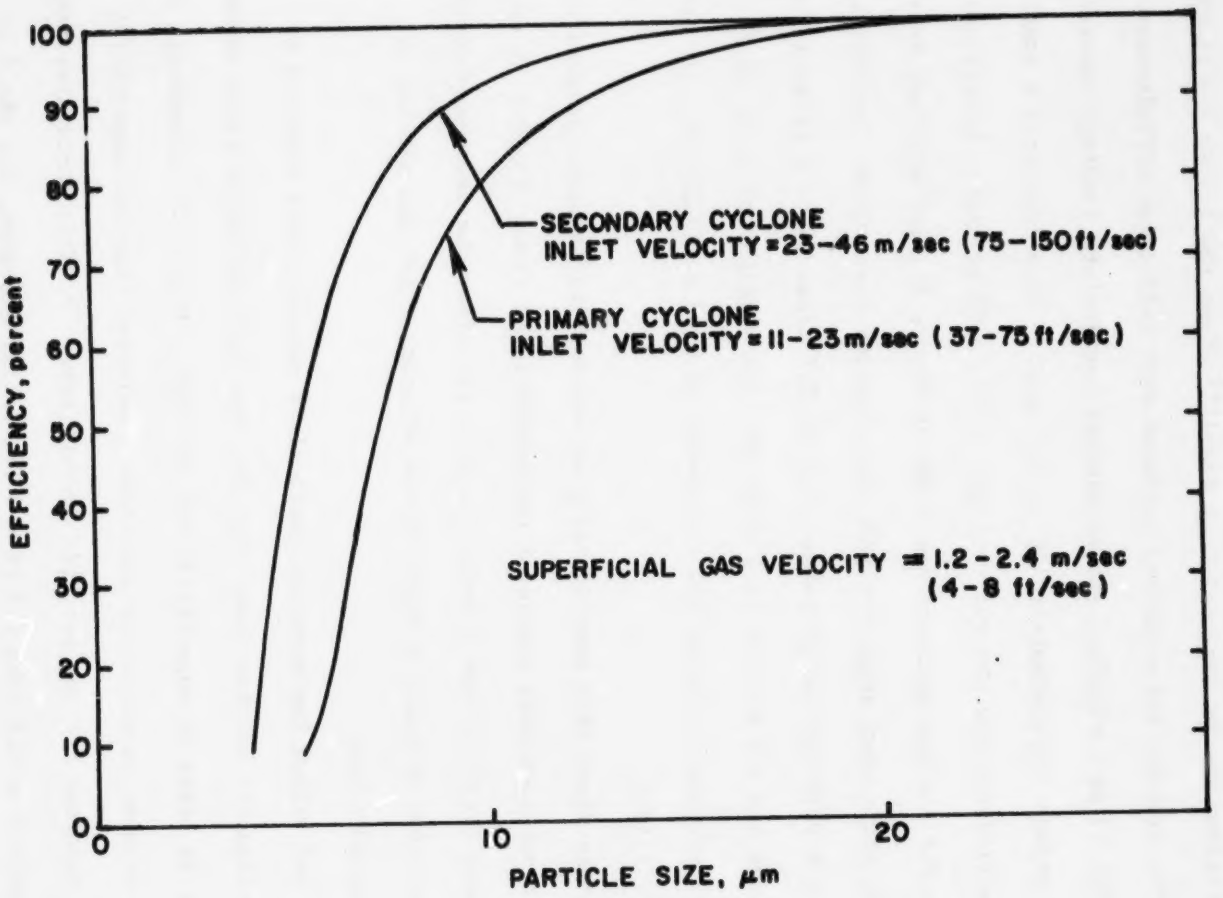

Figure 24. Fractional efficiency of the primary and secondary cyclones during experimentation in the NCB-CRE 36 in. × 18 in. test unit.

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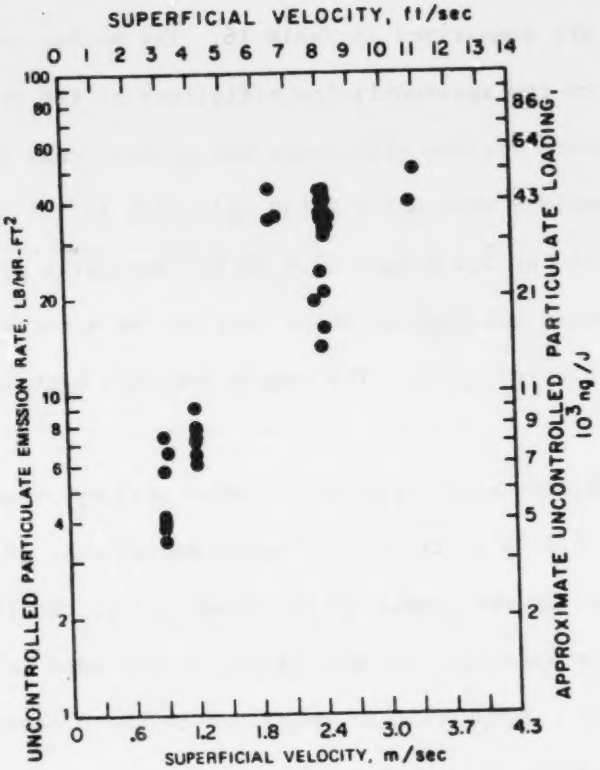

Figure 25. Uncontrolled particulate emission rate versus superficial velocity – Stone 7.[143] **(Reproduced with permission of EPRI.)**

In actual application, this relationship may not be so extreme (with respect to required removal efficiency in the final particulate collection device), because the primary and/or secondary cyclones will perform more efficiently in collecting particles of larger size elutriated at higher superfical velocities.

Babcock and Wilcox has reported results of particulate emission testing in their 6 ft × 6 ft unit.[144] Gas residence times were in the range of 0.4 to 0.6 sec with superficial velocities between 2.1 to 3.0 m/sec (7 to 10 ft/sec). Lowellville limestone was fed with a top size of 9,525 μm (3/8 in. × 0). Particle loadings were measured at the inlet and outlet of the primary cyclone using automatic duct traversing and isokinetic sampling (see Section 7.0).

The results are summarized in Table 16. The outlet loadings are fairly high, mainly due to the apparently low efficiency of the primary cyclone. In those instances where cyclone efficiency was greater than 80 or 85 percent, cyclone outlet loadings were below 2,150 ng/J (5.0 lb/10^6 Btu). One outlet loading was measured at 3,224 ng/J (7.5 lb/10^6 Btu) at a cyclone efficiency of 83 percent. However, the highest inlet loading (by a factor of 2) was measured during this test (18,900 ng/J). The reason for this high inlet loading is unclear.

Babcock and Wilcox also reported emission testing results from operation of their 0.91 m × 0.91 m (3 ft × 3 ft) experimental unit.[145] Particulate measurements were made in the combustor freeboard and in the flue prior to the wet scrubber inlet. An internal cyclonic cavity is included in the flue for primary particulate removal. Comparison of emissions in the freeboard (prior to the internal cyclonic cavity) and at the wet scrubber inlet illustrated a total collection efficiency ranging between 50 to 80 percent for the cyclone with an average capture of about 70 percent. This is below the capture efficiency of 85 to 90 percent normally cited as appropriate for primary particulate removal. Therefore, the particulate emission rates noted in Table 16 are higher than expected from an industrial FBC boiler utilizing a primary particulate removal device with efficiency of 85 to 90 percent. Other factors contributing to the high particle emissions are the low freeboard of the unit and the relatively high superficial velocities used during the testing, between 2.4 to 3.7 m/sec (8 to 12 ft/sec).

However, the data show how particle elutriation varies as a function of sorbent particle size and feed rate. During addition of Lowellville limestone with a top size of 6,350 μm, measured particulate rate after the cyclone was between 2,253 to 3,375 ng/J (5.24 to 7.85 lb/10^6 Btu). As limestone top size was decreased to 2,380 μm, the range in particulate rates measured at the wet scrubber inlet increased to 2,878 to 3,689 ng/J (6.74 to 8.58 lb/10^6 Btu). Dropping limestone size further to 1,000 μm top size, the range in particulate rate increased to 3,078 to 4,170 ng/J (7.16 to 9.70 lb/10^6 Btu). With pulverized limestone, the particulate rate was measured in the range of 5,434 to 7,825 ng/J (12.64 to 18.20 lb/10^6 Btu). Using pulverized Greer and Grove limestones and $Ca(OH)_2$ with a top size of 44 μm produced particulate rates of approximately 6,450 ng/J (15 lb/10^6 Btu) with a maximum of 15,215 ng/J (35.39 lb/10^6 Btu).

Again, the low freeboard of the B&W 3 ft × 3 ft unit, combined with the high gas velocity, undoubtedly contributed significantly to the high particulate emissions. The trend in envisioned commercial FBC units is to design with higher freeboard and lower superficial gas velocities.

2.5.4 Summary of Particulate Emission Data

The particulate summary table (Table 16 in Section 2.5.3.2) summarizes the particulate data presented in this subsection for atmospheric FBC systems. Emissions measured downstream of primary and secondary cyclones are specified along with associated removal efficiencies.

The cyclone outlet emissions measured from the B&W 6 ft × 6 ft test unit are slightly higher than would be expected in a commercial unit operating with a high primary cyclone efficiency. In most of these tests, primary cyclone efficiency was below 75 percent. When efficiency approached 85 percent, emissions generally fell below 2,150 ng/J (5.0 lb/10^6 Btu).

The Babcock and Wilcox emission data recorded at the inlet to the wet scrubber of the 3 ft × 3 ft unit are significantly higher than any of the data from other units due to the low freeboard on the 3 ft × 3 ft unit. However, good primary particulate control conditions were not noted during this experimentation. Considering the high fluidizing velocity, shallow bed, and primary collector design, the emissions measured in the Babcock and Wilcox 3 ft × 3 ft unit are essentially uncontrolled in comparison to other FBC units with deeper beds and better primary cyclone designs.

Particulate control requirements for AFBC should be similar to requirements for conventional boilers burning low sulfur coal. Use of a cyclone alone is not adequate to attain emission levels as stringent as 43 ng/J (0.1 lb/10^6 Btu) or lower. Demonstration of control equipment is necessary because very little data exist to support the capability of final particulate control devices applied to atmospheric FBC boilers.

Based on PER, NCB, and ANL data, it appears that dust loadings entering final control systems are in a range (0.5 to 5.0 lb/10^6 Btu) similar to emissions generated in conventional systems. Mass mean particle sizes entering the final control systems may be about 5 to 20 μm, depending upon the design of the cyclones. Therefore, with application of add-on equipment, any standard for conventional sources should also be supported by FBC. Future test programs to be conducted at Rivesville, West Virginia, Georgetown University, EXXON, the EPA-SATR test unit, and other sites will indicate performance capabilities of ESPs, fabric filters, and wet scrubbers used as final particulate control devices.

2.5.5 Impacts of Particle Control on Boiler Operation

It is not expected that use of add-on final particulate control systems will have any adverse impact on industrial FBC boiler operation.

2.5.6 Documentation

As summarized in Section 2.2.1.4, available source test data demonstrating the efficiency of final particulate control is very limited. Data presented here are based on studies conducted at ANL, PER, NCB, and Babcock and Wilcox.

2.6 PRESSURIZED FBC

Pressurized FBC boilers would only be used in industrial applications if the user had large electricity requirements and the system adequately fulfilled the specific cogeneration needs. Based on the stage of PFBC development, it is not anticipated that the typical industrial user would have sufficient need for electrical power (from a gas turbine) to warrant the increased capital cost and system complexity involved. Therefore, we have not considered pressurized fluidized-bed boilers in this report. Although specific larger industries might use pressurized technology we do not anticipate widespread application in the near future.

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3.0 CANDIDATES FOR BEST SYSTEM OF EMISSION REDUCTION

3.1 CRITERIA FOR SELECTION

The criteria used in selecting best systems of emission reduction are as follows:

- System Performance - Ideally, the technique chosen for any one pollutant should have the least possible impact on combustion or boiler efficiency, the least possible impact on system operability, and result in the least possible increase in the emissions of other pollutants from the system.
- Applicability - The best system of emission reduction should have a relatively wide applicability across the spectrum of boilers to be encountered in the industrial sector. It should not be especially sensitive to factors such as size, fuel type, load cycle, plant configuration, etc.
- Status of Development - The emission control technique should be forecasted as being available when emission control levels are set and incorporated into AFBC units as they find widespread commercial acceptance. It would be better if the techniques were available now, or at least in a prototype status. The best situation would be for the tecniques to be already available and successfully demonstrated.
- Cost - The system should be capable of meeting optional emission control levels without inordinate increases in capital or operating cost. Ideally, the "best system" would have the lowest cost of the options available.

For SO_2, the best system of emission reduction is the one which minimizes sorbent feed rates, and still attains high levels of control. The Ca/S molar feed ratio can be reduced with careful control of other operating conditions; most significantly, sorbent particle size and gas phase residence time. Reduction of sorbent requirements reduces not only the operating cost associated with purchase of fresh sorbent, but also reduces the cost and environmental impact of spent solids disposal. Electricity requirements are reduced and boiler efficiency is slightly increased.

Emission control techniques for SO_2 which were rejected either because of limited applicability or still tenuous technical development included pressurized fluidized-bed combustion, sorbent precalcination, sorbent regeneration, synthetic sorbents and sorbent catalysts. The use of SO_2 scrubbers on FBC flue gas was also not considered, due to concerns regarding performance and applicability.

For NO_x emissions, the best system of emission reduction appears to be capitalizing on the inherent combustion chemistry which occurs in fluidized-bed systems. The low temperature and the chemical kinetics of the system combine to provide relatively low NO_x emissions. For stringent NO_x control some care in the selection of design/operating conditions may be required. Control techniques which were not considered, due primarily to status of development, include substantive combustion modifications (such as two-stage combustion, flue gas recirculation, and ammonia/urea injection), and NO_x scrubbing.

It is expected that particulate emissions can be controlled using conventional particulate control technology which is currently available, with the best systems appearing to be fabric filtration or electrostatic precipitation for stringent or intermediate control, and multitube cyclones for moderate control. Neither fabric filtration nor electrostatic precipitators have yet been tested on commercial-sized FBC facilities, but pilot plant data have not suggested any unusual problems beyond those that would be encountered in a conventional boiler burning low sulfur coal.* By suitable design and operation of the particle control devices, it is anticipated that satisfactory performance can be achieved.

3.1.1 Selection of Optional Emission Control Levels

The optional emission control levels which will be addressed in selecting best systems are shown in Table 17. These ranges of optional emission control levels have been chosen because they are felt to represent attainable control levels using FBC boilers. The rationale for selection of these optional levels is discussed further in the following subsections. All conclusions are based on initial test results from prototype units, and from more extensive data compiled during operation of small FBC test units. In some cases, conclusions have been supplemented by current theory concerning the FBC process. They are subject to change when larger units come online and better data are available.

In the ensuing discussion of emission control technologies, candidate technologies are compared using three emission control levels labelled "moderate, intermediate, and stringent." These control levels were chosen only to encompass all candidate technologies and form bases for comparison of technologies for control of specific pollutants considering performance, costs, energy, and nonair environmental effects.

From these comparisons, candidate "best" technologies for control of individual pollutants are recommended for consideration in subsequent industrial boiler studies. These "best technology" recommendations do not consider combinations of technologies to remove more than one pollutant and have not undergone the detailed environmental, cost, and energy impact assessments necessary for regulatory action. Therefore, the levels of "moderate, intermediate, and stringent" and the recommendation of "best technology" for individual pollutants are not to be construed as indicative of the regulations that will be developed for industrial boilers. EPA will perform rigorous examination of several comprehensive regulatory options before any decisions are made regarding the standards for emissions from industrial boilers.

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*Several performance tests are scheduled at Georgetown University in early 1980.

TABLE 17. OPTIONAL LEVELS OF CONTROL TO BE SUPPORTED — ATMOSPHERIC FLUIDIZED-BED COMBUSTION OF COAL

Level of control	SO_2 % reduction	NO_x ng/J (lb/10^6 Btu)	Particulate ng/J (lb/10^6 Btu)
Stringent	90*	215 (0.5)	12.9 (0.03)
Intermediate	85*	258 (0.6)	43 (0.1)
Moderate	75*	301 (0.7)	107.5 (0.25)

*In addition to the % reduction, an upper limit of 516 ng/J (1.2 lb/10^6 Btu) applies in all cases. Furthermore, in no case are controls required to reduce emissions below 86 ng/J (0.2 lb/10^6 Btu).

3.1.2 Selection of SO_2 Emission Levels

3.1.2.1 Moderate Level of Control: 75 percent removal, 516 ng/J (1.2 lb/10^6 Btu) ceiling, 86 ng/J (0.2 lb/10^6 Btu) floor--

The moderate level of control can be supported by normal engineering application and operation of fluidized-bed combustion boilers. This degree of control has been consistently demonstrated by all investigators who have experimented with sorbent addition for SO_2 removal. Babcock and Wilcox Company compiled and reviewed available data on the operation of atmospheric fluidized-bed combustion (AFBC).[1] Considering 368 data points, the average Ca/S molar feed ratio was 2.2 with an attendant SO_2 reduction of 76 percent, demonstrating that the moderate level of control should be attainable on a routine basis.

3.1.2.2 Stringent Level of Control: 90 percent removal, 516 ng/J (1.2 lb/10^6 Btu) ceiling, 86 ng/J (0.2 lb/10^6 Btu) floor--

This high level of SO_2 reduction has not been widely demonstrated in AFBC experimentation to date, but theoretical projections by Westinghouse[2] and some experimental results indicate that 90 percent control is technically and economically achievable. The B&W 6 ft × 6 ft unit[3] and the B&W, Ltd., Renfrew[4] boiler have demonstrated greater than 90 percent SO_2 reduction at Ca/S ratios of 4 or less. Several other test units have achieved reductions as great as 90 percent, but only intermittently. These test data are shown in Section 7.0.

Westinghouse Research and Development has formulated an SO_2 removal model for FBC.[5] The model predicts SO_2 removal efficiencies based on sulfation rate constants measured in laboratory thermogravimetric analysis apparatus, and considers sorbent parameters and FBC operating conditions. The important factors considered in the model are Ca/S molar feed ratio, gas phase residence time and sorbent particle size. The FBC conditions suggested by the Westinghouse model for effective SO_2 control (i.e., our definition of the "best system") are a gas phase residence time of 0.67 sec (superficial gas velocity of 1.8 m/sec and expanded bed depth of 1.2 m) and average inbed sorbent particle size of 500 μm. Most current FBC designs incorporate shorter gas residence times and larger particle size;* however, these conditions are well within the range that has been considered in previous studies by others (notably NCB, FluiDyne,

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*Refer to Table 20. Although differences in design/operating conditions exist between current designs and those conditions recommended here for the "best system" of SO_2 control, the differences are not great and could be adopted with only minor modifications in current boiler designs.

B&W) and are felt to be useable in many existing designs without major redesign difficulty. At these conditions, 90 percent SO_2 control should be achieved at Ca/S ratios between 2.5 and 4.0, based on the model, and some experimental results. No higher control level was considered because extrapolation of the existing data base is too uncertain. It is important to emphasize that possible capital cost increases associated with going to "best system" conditions should be offset by reduced operating costs and possible capital savings in other areas of the system (see Section 4.3.4).

One technical uncertainty which exists regarding SO_2 control is whether overbed solids feeding allows for "best system" gas residence time. In applicable experimentation by FluiDyne in their 1.5 ft × 1.5 ft unit,[6] they found equivalent high levels of SO_2 reduction (>90 percent) with underbed or overbed feed as long as primary recycle of bed carryover was practiced. This is discussed further in Sections 3.2.1.2 and 7.5.6.

The Westinghouse model has been reasonably well confirmed at lower levels of SO_2 control (85 percent and less) by comparison with experimental results which are available in the literature, as shown in Section 7.0, Subsection 7.7. Since experimental data at 90 percent SO_2 reduction are limited, the model cannot yet be reliably confirmed at this degree of control. However, based on the apparent validity of the model at lower desulfurization levels (85 percent and less) and the actual data which do not exist, we conclude that 90 percent reduction will be achievable in industrial AFBC boilers at Ca/S ratios between 2.5 to 4.0.*

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*The Westinghouse model assumes uniform SO_2 generation throughout the depth of the bed. In underbed feed systems, where SO_2 may be preferentially formed near the bottom of the bed, the Westinghouse model may predict less efficient SO_2 removal than actually achievable.

Based on the results of the cost analysis presented in Section 4.0, 90 percent SO_2 control can be achieved with little additional economic impact compared to 75 or 85 percent control. In fact, for high sulfur coal, the maximum incremental cost of going from 75 to 90 percent desulfurization, is about \$0.30/10^6 Btu output. This is about 5 percent of total AFBC system cost. The cost penalty is insignificant when low sulfur coals are considered. These results are confirmed by independent estimates which Westinghouse has made for industrial AFBC boilers (see Section 4.0, and Appendix D).[7] The cost of SO_2 control is more sensitive to sorbent reactivity than to the degree of control required when levels greater than 75 percent reduction are considered.

Energy and environmental impacts are also only slightly increased if 90 percent SO_2 reduction is employed, compared to moderate control levels. The energy analysis in Section 5.0 indicates that AFBC boiler efficiency is comparable to and potentially greater than conventional stoker technology, even when the SO_2 controlled AFBC case is compared to the conventional boiler with no SO_2 control. With use of "best system" design/operating conditions, energy efficiency is not significantly impacted by adding sorbent to the bed. The major energy impact of either FBC or conventional technology is flue gas heat loss which overshadows the impact of SO_2 control. Conventional pulverized coal (PC) technology has generally higher boiler efficiency than AFBC due to better combustion efficiency. However, if coal drying is necessary in the PC case, and not necessary for AFBC (assuming overbed feeding with primary recycle), then AFBC boiler efficiency can be comparable to PC technology.

The only environmental impact which is increased by going to 90 percent SO_2 control, is solid waste disposal. Particulate control capability should not suffer; i.e., the optional levels considered in this study can still be met. NO_x emissions are unaffected and may even be reduced using the "best system" of SO_2 control because longer gas phase residence time allows for further chemical reduction of NO. Solid waste quantities are greater at 90 percent control than at lower levels, but again, if sorbent reactivity is reasonable, it should not be an overwhelming problem.

3.1.2.3 Intermediate Level of Control: 85 percent removal--

An intermediate SO_2 control level of 85 percent has been chosen because it represents about the most stringent level of control which has been consistently demonstrated by most investigators (those who have used sufficient sorbent quantities and appropriate operating conditions). Moreover, modeling studies project that, with suitable FBC design (appropriate gas residence times and sorbent particle sizes), this degree of control can be achieved at moderate sorbent feed rates (Ca/S = 2 to 3.5, with sorbents of reasonable reactivity).[8]

3.1.2.4 Upper and Lower Limits of Control Levels: 516 ng/J ($1.2\ lb/10^6$ Btu) upper, 86 ng/J ($0.2\ lb/10^6$ Btu) lower--

These levels of emissions are being specified to allow flexibility in burning a variety of fuels with a wide range of sulfur contents. The lower limit of 86 ng/J ($0.2\ lb/10^6$ Btu) is proposed to allow for burning of low sulfur fuels without requiring excessive percentage reductions of SO_2. The upper limit of 516 ng/J ($1.2\ lb/10^6$ Btu) assures that the optional levels considered are not more lenient than standards previously established for electric utility boilers. Table 18 shows the various levels of control for several fuels with sulfur contents ranging from 0.5 to 3.5 percent. The table indicates the limiting factor for each level of control. Notice that for a fuel containing 1.0 percent sulfur or less, the floor of 86 ng/J can be met by less than 90 percent reduction of SO_2; and that for fuels containing 3.5 percent sulfur or greater, SO_2 reduction must be greater than 75 percent, the proposed moderate level of control, to insure that emissions do not exceed the ceiling of 516 ng/J (1.2 lb/10^6 Btu).

TABLE 18. SO_2 CONTROL LEVELS FOR FUELS OF VARYING SULFUR CONTENT

% Sulfur	Uncontrolled SO_2 emissions[*] ng/J (lb/10^6 Btu)	Control level	Required % SO_2 reduction	Controlled SO_2 emissions ng/J (lb/10^6 Btu)
0.5	344 (0.8)	Stringent	75	86 (0.20)[†]
1.5	1,032 (2.4)		90[†]	103 (0.24)
3.5	2,365 (5.5)		90[†]	236 (0.55)
0.5	344 (0.8)	Moderate	75[†]	86 (0.20)[†]
1.5	1,032 (2.4)		75[†]	258 (0.60)
3.5	2,365 (5.5)		78	516 (1.20)[†]
0.5	344 (0.8)	Intermediate	75	86 (0.20)[†]
1.5	1,032 (2.4)		85[†]	155 (0.36)
3.5	2,365 (5.5)		85[†]	357 (0.83)

[*] Coal HHV = 28,000 kJ/kg.

[†] Limiting level of control.

3.1.3 Selection of NO_x Emission Levels

The mechanisms by which NO_x is formed in FBC, and by which NO_x can be controlled, are not understood as well as in the case of SO_2. NO_x emissions tend to be low in FBC because of the prevailing chemistry within the bed. Past work has involved primarily just the monitoring of NO_x emissions from FBC units, with some effort to explore the impact on emission of some key variables. Concentrated efforts to model and reduce emissions of NO_x from FBC are just beginning.

3.1.3.1 Moderate Level of Control: 301 ng/J (0.7 lb/10^6 Btu)--

All data from the larger FBC test units have been consistently below 301 ng/J (see Figure 27 in Subsection 3.2.2), except at temperatures which are higher than envisioned for typical AFBC operation (>1000°C). Despite its small size, most data from the ANL 6 in. bench-scale unit are below this level. Therefore, this level should be supported by FBC boilers under normal operating conditions.

3.1.3.2 Stringent Level of Control: 215 ng/J (0.5 lb/10^6 Btu)--

The average NO_x emission rate observed in past experimentation at typical AFBC operating temperatures is in the range of 375 ppm NO_x, which corresponds to 215 ng/J NO_x (see Figure 27). In addition, data from the large units which have come on stream recently (Renfrew, and the EPRI/B&W 6 ft × 6 ft unit) are consistently less than 215 ng/J (generally between 165 and 215 ng/J, or 0.4 to 0.5 lb/10^6 Btu). This level (215 ng/J) has thus been designated as achievable for a stringent level of control; it is considered to be the lowest level that a manufacturer can guarantee at this time. Although emissions of less than 215 ng/J (0.50 lb/10^6 Btu) have been observed fairly frequently, the role of the factors which control NO_x formation and decomposition in the bed (such as fuel nitrogen, gas residence time, excess air, and temperature) is not sufficiently well understood; the correlation between NO_x emissions and the variables which have been studied to date, does not appear to be significant based on existing data.[9-11] Therefore, control of these parameters cannot at this time be relied upon to ensure NO_x emissions below 215 ng/J and, in fact, further data from the large FBC units would be desirable to ensure that 215 ng/J itself would be reliably achievable on a 24-hour average.

Experimental studies are in progress at MIT specifically for characterization of NO_x formation and control in FBC.[12,13] The stringent level considered here has been consistently attained in their pilot-scale unit.

Some conventional boiler controls may be applicable for maintenance of reduced NO_x emissions from FBC systems. The use of low excess air levels and two-stage combustion may aid in reducing NO_x emissions reliably. However, combustion modifications for FBC have not yet been extensively studied. Such modifications could impact materials corrosion, combustion efficiency and emissions of other pollutants. Further research and development is required on FBC combustion modifications, although such modifications are not considered available control technology for the purpose of this document.

3.1.3.3 Intermediate Level of Control--

In the temperature range of interest (815° to 870°C) for primary FBC combustion cells, virtually all of the available data from large AFBC units (500 lb coal/hr and larger) are below 260 ng/J. Even most of the data from smaller experimental units are below this level. Therefore, 260 ng/J has been selected as the intermediate level of control.

3.1.4 Selection of Particulate Emission Levels

It is expected that a primary cyclone will be used as an integral part of first generation atmospheric fluidized-bed combustion industrial boilers. The purpose of the primary cyclone is to recycle elutriated sorbent to increase sorbent/SO_2 contact time, recycle unburned carbon to the combustor, prevent fire hazards in the downstream flue gas ducting, and decrease the particulate loading to the final particulate control device. Primary cyclone efficiency should be in the range of 80 to 90 percent, depending on FBC operating parameters and cyclone design.

Particulate emissions following the primary cyclone in coal-fired atmospheric FBC systems and final particulate reduction necessary to meet stringent, intermediate, or moderate standards are shown in Table 19.

TABLE 19. REQUIRED PARTICULATE CONTROL EFFICIENCIES FOLLOWING THE PRIMARY CYCLONE IN COAL-FIRED ATMOSPHERIC FBC SYSTEMS

Fuel and boiler capacity MWt (10^6 Btu/hr)	Particulate emission following primary cyclone ng/J (lb/10^6 Btu)	Particle size average MMD μm	Level of emission control and efficiency of final particulate control device required to achieve that level ng/J (lb/10^6 Btu)		
			Stringent 12.9 (0.03)	Intermediate 43 (0.10)	Moderate 107.5 (0.25)
Coal					
8.8 - 58.6 (30 - 200)	215 - 2150 (0.5 - 5.0)	5 - 20	94 - 99.4	80 - 98	50 - 95

The emission range of 215 to 2,150 ng/J (0.5 to 5.0 lb/10^6 Btu) is based on particulate data shown in Sections 7.0 and 2.0. Pope, Evans, and Robbins,[14] Argonne,[15] and NCB[16] have measured emissions after the primary cyclone between 215 to 960 ng/J (0.5 to 2.0 lb/10^6 Btu). Babcock and Wilcox[17,18] has measured higher emissions from their 6 ft × 6 ft and 3 ft × 3 ft units, but in cases where outlet loadings were greater than 2,150 ng/J (5.0 lb/10^6 Btu), primary collection efficiencies were poor. The 3 ft × 3 ft unit is not representative because a low efficiency cyclonic cavity was used for primary control. In addition, freeboard height was low and a shallow bed was used. The B&W 6 ft × 6 ft unit indicated higher outlet loadings than 2,150 ng/J (5.0 lb/10^6 Btu) mainly when primary cyclone efficiency fell below 75 percent. Therefore, the upper limit on uncontrolled particle emissions (i.e., the outlet from the primary cyclone) is reported here as 2,150 ng/J (5.0 lb/10^6 Btu). The mass mean particle size in the primary cyclone outlet, based on available data, appears to be in the range of 5 to 20 μm.

Although final particulate control has not been thoroughly demonstrated in AFBC systems to date, it is expected that final particulate control in industrial AFBC boilers will be as effective as and similar to, conventional systems burning low sulfur coal. Conventional particle control technology, suitably designed and operated for FBC applications, should provide the necessary control.

3.1.4.1 Moderate Level of Control: 107.5 ng/J (0.25 lb/10^6 Btu)--

Due to the wide range in expected particulate loadings to the final control device, the control efficiency required to meet a moderate particulate level of 107.5 ng/J (0.25 lb/10^6 Btu) ranges from 50 to 95 percent. The moderate level was selected because this range is well within the capabilities of conventional particle control technology. With a mass median particle diameter of 10 μm or greater, conventional multitube cyclones should be capable of providing 50 to 80 percent removal efficiency. If either lower mass median diameters exist (5 to 10 μm) or greater control efficiencies (80 to 95 percent) are required, use of other control devices such as ESPs, or fabric filters, will be necessary.*

3.1.4.2 Stringent Level of Control: 12.9 ng/J (0.03 lb/10^6 Btu)--

Stringent control requires final collection efficiencies ranging between 94 to 99.4 percent. Although this level of control has not been demonstrated in AFBC systems, it was selected because it is anticipated that it can be supported using fabric filters or possibly ESPs, based on performance demonstrated in conventional boilers.[19]

3.1.4.3 Intermediate Level of Control: 43 ng/J (0.10 lb/10^6 Btu)--

This level has been established to demonstrate the various impacts associated at midrange control level. At least in conventional boiler installations, it has been demonstrated as a critical value above which significant costs and energy penalties may occur.

Final particle removal efficiencies between 80 and 98 percent are required to attain an intermediate particulate control level. This range of control should be achievable using fabric filters or ESPs. Multitube cyclones may also be applicable depending on actual particle sizes and efficiency requirements.

3.1.5 Impact of Averaging Time

The time period over which emissions are averaged may influence FBC operating requirements to meet optional control levels. In the case of SO_2, Ca/S may vary with time due to changes in coal sulfur, sorbent reactivity, boiler loading or other conditions. These effects have not been rigorously explored in experimentation to date. More testing is required for longer time periods to determine whether a safety factor in Ca/S requirements is necessary if averaging times of 24 hours or longer are considered. Potential impacts on NO_x and particulate emission levels must also be characterized.

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*If a sliding scale based on boiler size is used for particulate control such that smaller boilers have less stringent control demands, multiclones may be the most cost-effective technique for smaller units.

3.2 BEST CONTROL SYSTEM FOR COAL-FIRED BOILERS

The following discussion specifies the data on which the choices of best control techniques were made. The discussion follows each of the specific pollutants, namely SO_2, NO_x, and particulates. In many cases, supporting data from other sections of the report are referenced and not reproduced here. Controls for coal-fired boilers are emphasized in this report.

Since data from commercially-operating AFBC units are not available, the selection of "best systems" is necessarily made based upon laboratory and pilot plant data, and upon projections prepared using these data and engineering principles.

3.2.1 SO_2 Emissions

3.2.1.1 Factors Affecting SO_2 Control--

The primary factors influencing SO_2 control are the following:

- Calcium to sulfur molar feed ratio
- Type of limestone
- Particle size
- Gas phase residence time

The Ca/S molar feed ratio is usually varied to control the level of SO_2 emissions from fluidized-bed combustion. In order to maximize the overall efficiency and performance of an FBC system, at a specific level of SO_2 control, the Ca/S ratio must be minimized to reduce sorbent feed quantities and to minimize waste disposal problems. Among the calcium-based sorbents which have been used in FBC systems, there are a wide range of reactivities. However, it is not likely that a sorbent will be chosen solely on the basis of its reactivity, but rather, will be selected primarily on the basis of the proximity of the quarry to the FBC facility. Thus, the particle size and gas phase residence time become the important factors in obtaining the best results. Reducing particle size and increasing gas phase residence time can increase calcium utilization and allow for significantly lower Ca/S ratios to support a specific level of SO_2 reduction.[20] In some instances, these modifications would require some redesign of current FBC systems.

Particle size and residence time have historically been set by FBC designers based on considerations other than SO_2 control. The effort has been to make the boiler as small as possible to allow for shop fabrication of boilers of larger capacity than traditionally possible by increasing velocity (decreasing the residence time) and hence, also increasing the required sorbent particle size. Much of the experimental work to date has not been conducted at residence times felt to approach the optimum for SO_2 control (0.67 sec or greater). In addition, some designs (especially overbed coal feed designs and inherently shallow-bed designs) may not readily lend themselves to adjustment of gas residence time. However, our estimates indicate that, although increased gas residence time will result in somewhat larger boilers and possibly higher boiler cost, this higher cost will be more than offset by the reduced sorbent requirements. Thus, reasonable increases in gas phase residence time and correspondent decreases in particle size are presented in this report as the best system of SO_2 control for AFBC.

The optimal values for sorbent particle size and gas phase residence time cannot be specifically defined based on currently available information; however, an estimate of close to optimal values can be made.[21] It is not clear whether technical or economic factors will limit the degree to which sorbent particle size can be lowered or gas phase residence time can be increased. Increased gas residence time and decreased sorbent particle size may increase boiler costs at the same time they decrease sorbent requirements and cost. For any specific site and sorbent, there may be an economically-determined optimum combination of residence time and particle size which minimize cost of steam from the boiler. On the other hand, if the economics continue to look attractive as the terminal particle velocity falls below the minimum fluidization velocity, technical factors, rather than economic, could become the limiting concern. Specifically, using very fine particle sizes of 100 μm or less could alter fluidization needs, requiring high recycle or "fast" fluidization. A design for such a fast bed currently exists[22] but it is still under development. Additionally, there may be a point of diminishing returns in SO_2 control with extremely small particle sizes or long gas residence times.

The Westinghouse calculations suggest that gas phase residence times in the neighborhood of 0.67 sec, and sorbent particle sizes in the neighborhood of 500 μm should be suitable for effective SO_2 removal at reduced sorbent feed rates. (The 0.67 sec residence time results using a 1.2 m deep bed and a 1.8 m/sec gas velocity.) These are the conditions which will be considered for the "best system" of SO_2 control in this report. However, this particular combination of conditions will not necessarily be the economic optimum for all AFBC systems; the true optimum will vary from one specific case to another, depending upon the specific site and sorbent characteristics. (For example, in one case, a reduced gas residence time may be desirable in order to result in a boiler small enough for shop fabrication.) It is felt, however, that this combination of conditions will be sufficiently representative of the optimum for all cases, so that it is used in this report to indicate the performance and cost of "best" SO_2 control systems. The smaller particle size (500 µm) is suggested assuming that the primary cyclone catch will be recycled. If packaged FBC units (with low freeboard) did not employ recycle, coarser (1,000 µm) sorbent might be needed to maintain the bed, thus increasing the Ca/S requirement. The residence time and particle size chosen represent a breakpoint in the relationship of gas residence time and Ca/S requirements and particle size and Ca/S requirements, according to Westinghouse data.[23]

3.2.1.2 Selected Design/Operating Conditions for the "Best System" of SO_2 Control--

Based on the preceding discussion and other considerations mentioned below, "best system" design/operating conditions for SO_2 control in FBC are represented by the following values:

- Bed depth — 1.2 m (4 ft)
- Superficial gas velocity — 1.8 m/sec (6 ft/sec)
- Gas phase residence time[*] — 0.67 sec
- Sorbent particle size — 500 µm (32 mesh) inbed surface average[†]
- Coal and sorbent feed — Inbed or abovebed
- Primary recycle — Yes, for either feed orientation
- Bed temperature — 843°C (1550°F)
- Excess air — 20 percent

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[*] Estimated by dividing bed depth by superficial gas velocity.

[†] A 500 µm surface average is roughly equal to a mass average particle size between 600 to 700 µm, depending on the actual particle size distribution. Theoretically, at 1.8 m/sec (6.0 ft/sec) fluidizing velocity, surface average particle sizes between 350 to 1500 µm are suitable for operation, allowing for fluidization without significant sorbent loss through entrainment (assuming use of primary recycle). Actual particle distribution and combustor design would affect this range to some extent.

To date, the majority of experimental FBC units have operated with inbed coal and limestone feed during testing. This allows for SO_2 formation near the bottom of the bed and provides the maximum residence time for SO_2 to react with CaO, within the designated design/operating conditions of the unit.

One set of experiments has been conducted by FluiDyne in their 1.5 ft × 1.5 ft unit to assess the effect of solids feed orientation on desulfurization efficiency. The results of this testing are detailed in Section 7.0 of this report. The data indicate that equivalent desulfurization levels can be obtained with inbed or abovebed feed as long as primary recycle is practiced. At a Ca/S molar feed ratio of 3.0 (using limestone), 94 percent SO_2 reduction efficiency was obtained regardless of feed orientation, using primary recycle in both cases (see Figure 59). Although the supporting data are limited in number, and the unit tested was small, it is concluded for the purpose of this study that abovebed solids feed is applicable for "best system" SO_2 control in FBC. If in actuality, higher Ca/S ratios are required with overbed feed systems in comparison to the average values shown in the next subsection (see Table 20), it is believed that the added operating cost of additional sorbent purchase is within the accuracy band of total annual FBC system cost estimated in this report. In the event that an FBC customer were to purchase an FBC system using underbed feed to minimize sorbent requirements (if higher Ca/S ratios were deemed necessary with overbed feed) the resultant economics should also fall within the specified accuracy bands in Section 4.0.

A temperature of 843°C (1550°F) was selected because in experimentation performed to date, peak SO_2 removal has been found in the temperature range of 816° to 871°C (1500° to 1600°F). The excess air rate of 20 percent has been commonly used in past experimentation. A higher rate might aid SO_2 reduction, but could increase NO_x formation and decrease boiler efficiency.

3.2.1.3 Ca/S Requirements for the "Best System" of SO_2 Control Based on Experimental Test Data--

Table 20 shows the required Ca/S molar feed ratios found by investigators using sorbent particle sizes and gas phase residence times close to those suggested here for "best systems." These Ca/S ratios were interpolated from curves fitting the actual data points (see Section 7.0). The ranges noted at the bottom of Table 20 are used throughout this report as the required Ca/S ratios when "best system" design/operating conditions are considered.

Judging from the data in Table 20, the Westinghouse model projections are good estimates of performance which can be expected from AFBC units operating at or near "best system" conditions (see Section 7.0 for further comparisons).

Figure 26 is a summary of experimental SO_2 reduction measurements made in bench- and pilot-scale units operating at a wide range of conditions, including some conditions different from the noted "best system" conditions. The range of Ca/S ratios used to determine "best system" performance and cost at the optional control levels (from Table 20) are shown by the straight lines between 56 and 90 percent SO_2 reduction. These limits represent high and low sorbent reactivity. Limestone 1359 (Grove limestone) was used as the index of low sorbent reactivity, and limestone 18, and U.K. limestone, were used as the index of high reactivity. The figure illustrates that the majority of experimental data, including data from experimentation conducted at other than "best system" conditions, fall within the brackets of performance for the range of reactivity considered here. Most of the data below the line of low sorbent reactivity were obtained from two units, the B&W 3 ft × 3 ft unit and the PER-FBM unit. The B&W 3 ft × 3 ft[43] has a shallow bed and low freeboard which reduce the time available for the gas/solid reaction of the SO_2 and CaO, thus reducing the SO_2 capture efficiency. The PER-FBM data[44] were generated using

TABLE 20. REQUIRED Ca/S MOLAR FEED RATIOS FOR BEST SO_2 CONTROL BASED ON EXPERIMENTAL DATA

Source	Temperature (°C) (°F)	Gas phase residence time sec	Sorbent-reactivity H, M, L	Size μm	Ca/S needed to maintain control level						Reference and test ID
					75%	78.7%	83.2%	83.9%	85%	90%	
ANL	840 - 870 (1550 - 1600)	0.67	Limestone 1359 L	AVE 25	2.4	2.7	3.1	3.2	3.4	4.2	ANL-CEN-ES-1001[24] ANL-CEN-ES-1002[25] TESTS SA-1, SACC-5, SACC-6, SACC-9, SA-2
ANL	840 - 870 (1550 - 1600)	0.67 - 0.70	Limestone 1359 L	177 × 0	2.5	2.7	2.9	3.0	3.1	3.6	ANL-CEN-ES- 1001[26] ANL-CEN-ES-1002[27] TESTS SA-3, SA-4, BC-1, BC-6
ANL	870 (1600)	0.67	Limestone 1359 calcined H	AVE 25	2.0	2.0	2.1	2.1	2.2	2.3	ANL-CEN-ES-1001[28] SACC-1, SACC-4
ANL	870 (1600)	0.5 - 0.7	Limestone 1359 L	AVE 490 - 630	2.1	2.5	2.7	2.8	3.0	3.5	Paper by Vogel at Third International[29] Conference on FBC
NCB	850 (1560)	0.58	Limestone 18 H	AVE 210	1.9	2.0	2.3	2.4	2.5	3.1	PB-210-673[30] NCB September 1971
NCB	800 - 850 (1470 -1560)	0.5	Dolomite 1337 H	AVE 100-125	2.6	3.0	3.3	3.3	3.4	3.8	PB-210-673[31] NCB September 1971 p. 23, Task I, Test 4
NCB	750 - 850 (1380 - 1560)	1.86	Dolomite 1337 H	AVE 100-125	1.8	1.9	2.2	2.3	2.3	2.6	PB-210-673[32] NCB September 1971 p. 23, Task I, Test 4
NCB	800 - 850 (1470 - 1560)	0.5	Limestone 18 H	AVE 210	2.1	2.3	2.7	2.7	2.8	3.2	PB-210-673[33] NCB September 1971 p. 20, Task I, Test 1.2, 1.3, 2, 5

(continued)

TABLE 20 (continued).

Source	Temperature (°C) (°F)	Gas phase residence time sec	Sorbent - reactivity H, M, L	Size μm	Ca/S needed to maintain control level						Reference and test ID
					75%	78.7%	83.2%	83.9%	85%	90%	
NCB	800 - 850 (1470 - 1560)	0.67	Limestone 18 H	AVE 210	1.8	1.9	2.2	2.3	2.3	2.6	PB-210-673[34] NCB September 1971 p.20, Task I, Test 1.2, 1.3, 2, 5
NCB	800 (1470)	0.67	U.K. Limestone H	AVE 300-400	1.6	1.8	2.0	2.0	2.1	2.4	PB-210-673[35] NCB September 1971 p. 57, Task V
NCB	800 (1470)	0.67	Limestone 1359	AVE 210	2.8	3.0	3.4	3.5	3.5	3.8	PB-210-673[36] p. 58, Test V
	800 (1470)	1.00	Limestone 1359	AVE 210	2.3	2.4	2.7	2.7	2.8	3.3	PB-210-673[37] p. 58, Test V
	800 (1470)	0.67	Limestone 1359	125 × 0	2.0	2.3	2.7	2.7	2.8	3.5	PB-210-673[38] p. 58, Test V
NCB	800 (1470)	0.67	Limestone 18	AVE 210	1.8	1.9	2.1	2.2	2.2	2.6	PB-210-673[39] p. 88
ANL	800 (1470)	0.67	U.K. Limestone	NR	3.2	3.4	3.6	3.7	3.8	4.2	PB-210-673[40] p. 90
ANL	800 (1470)	0.67	Limestone 1359	NR	2.7	3.0	3.2	3.3	3.4	3.8	PB-210-673[41] p. 90
NCB	800 (1470)	0.67	Limestone 1359	NR	2.7	3.0	3.2	3.3	3.4	3.8	PB-210-673[42] p. 90
Range of Data											
High	870	1.86	Low	25	1.6	1.8	2.0	2.0	2.1	2.3	
Low	750	0.5	High	650	3.2	3.4	3.6	3.7	3.8	4.2	
Average					2.2	2.5	2.7	2.8	2.9	3.3	

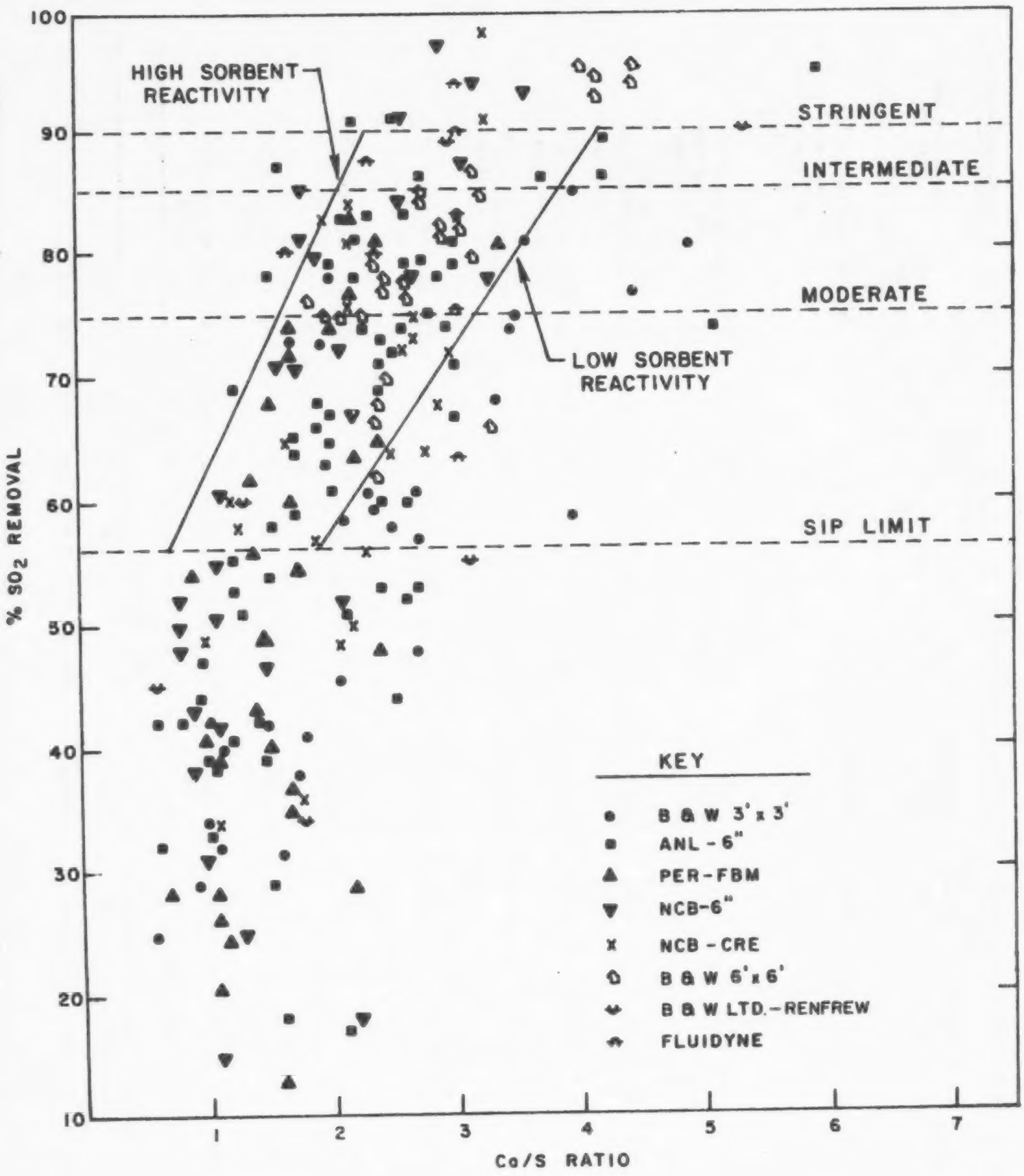

Figure 26. Summary of SO_2 reduction data measured in AFBC experimentation.

gas phase residence times as low as 0.2 sec, which, as in the case of the B&W 3 ft × 3 ft unit, does not give sufficient time for an efficient SO_2/CaO reaction. The graph also illustrates that all of the control levels under consideration in this study have been demonstrated in past testing using Ca/S ratios which are within a practical range. Further testing in larger units is required to confirm Ca/S needs at high levels (≥90 percent) of desulfurization.

3.2.1.4 Capability of Available FBC Systems Versus "Best Systems"--

Currently, there are several manufacturers offering FBC boilers on a commercial basis (see Table 9), but only limited sales have been documented. Other vendors will respond to a request for an FBC boiler but are not actively marketing units yet.

The design/operating conditions of "commercially-offered" FBC units are listed in Table 21 and are based on the larger experimental and demonstration units currently operating or in design (all but the CE/Great Lakes unit are currently in operation). All of the designs listed are representative of operating conditions that would be specified in commercial units. However, these conditions would vary on a site-specific basis.

The Westinghouse SO_2 removal model was used to project Ca/S ratios required for the "commercially-offered" boilers to meet the optional control levels under consideration. The resulting values are shown in Table 22. The sorbent requirements shown for the commercially-offered systems assume an average inbed sorbent particle size of 1,000 μm (surface mean) although the actual dimension may be different from this. This assumption was made for two reasons: (1) no documentation of actual inbed average sorbent size is provided by the vendors; and (2) most of the available sulfation rate data from Westinghouse are for particle sizes between 1,000 to 1,200 μm. The relationship

TABLE 21. COMMERCIALLY-OFFERED AFBC-INDUSTRIAL BOILERS — KEY FEATURES AFFECTING EMISSION CONTROL

Key Design/Operating Conditions	Foster-Wheeler Georgetown Design	Babcock & Wilcox, U.S. Alliance, Ohio 6'x6' Design	Combustion Engineering Great Lakes Design	Johnston Boiler Demonstration Plant Design	FluiDyne 40"x64" Vertical Slice Combustor Design [*]	B&W, Ltd. Renfrew Design	O. Mustad and Sons Enköping Design	Foster-Wheeler Rivesville Design
Reference Boiler Size	45,400 kg/hr steam		22,700 kg/hr steam			12 MW_t	25 MW_t	88 MW_t
Feeder type	Overbed	Underbed	Underbed	Overbed	Underbed	Underbed	Overbed	Underbed
Expanded bed depth, m (ft)	1.37 (4.5)	1.22 (4.0)	0.91 (3.0)	0.83 (2.7)	1.07-1.19 (3.5-3.9)	0.8-0.91 (2.6-3)	Slumped 0.25 (0.8)	1.2 (4)
Gas velocity, m/sec (ft/sec)	2.44 (8)	2.44 (8)	2.13 (7)	1.83 (6)	0.61-1.83 (2-6.0)	2.44 (8)	2.5 (8.2)	3.6 (12)
Approx. gas residence time, (sec)	0.56	0.50	0.43	0.44	0.58-2.0	0.35	—	0.33
Primary recycle	Yes	Yes	Yes	Yes	Yes	Yes	No	Yes
Sorbent type[†]	Greer, Grove	—	—	—	Dolomite	—	Sala dolomite	Carbon limestone
Ca/S ratio and % removal[‡]	3; 90	4; 90	3; 85	2; 75-95	—	3.0-5.5; 90	1.5;75	—
Sorbent size, μm (in. or mesh)	<4760 (4 mesh)	<9510 (3/8")	(¼"x0)	100% <2380 μm (8 mesh) 85% >1190 μm (16 mesh)	<6350 μm or <2380 μm	—	500-3000	1/8" x 16 mesh
Bed temperature, °C(°F)	868 (1594)	843 (1550)	843 (1550)	843 (1550)	718-796[§] (1325-1465)	849 (1560)	849 (1560)	816-843 (1500-1550)
Excess air, %	20	21	20	25	30-130	20	10[**]	15-20

[*] Although this unit is smaller than the others listed, the design/operating conditions are representative of FluiDyne's commercially-offered design, up to an air heating rate equivalent to 18,000 kg/hr steam.

[†] Sorbent type may vary significantly based on the geographic location of the installation.

[‡] As claimed by vendor.

[§] Higher temperature may be used in commercial units.

[**] Two-stage combustion.

TABLE 22. PROJECTED Ca/S RATIOS REQUIRED FOR "COMMERCIALLY-OFFERED" FBC BOILER SYSTEMS BASED ON THE WESTINGHOUSE MODEL

Coal type and SO_2 control level	% SO_2 removal	Greer limestone - high reactivity			Grove limestone - low reactivity		Tymochtee dolomite - high reactivity	Western 90% Cal limestone - high reactivity				Bussen Quarry limestone - medium reactivity					Menlo Quarry limestone - low reactivity			
		FW Georgetown design	FW Rivesville design	Best* system	FW Georgetown design	Best* system	FluiDyne 40" × 64" test unit	B&W (U.S.)	Combustion Engineering or Johnston Boiler Co.	B&W Ltd.	Best* system	FluiDyne 40" × 64" test unit	B&W (U.S.)	Combustion Engineering or Johnston Boiler Co.	B&W Ltd.	Best* system	B&W (U.S.)	Combustion Engineering or Johnston Boiler Co.	B&W Ltd.	Best* system
Eastern high sulfur																				
Stringent	90	5.29	5.63	2.85	>10	4.20	2.05	4.58	4.98	5.69	2.83	5.23	6.70	7.38	8.62	3.41	>10	>10	>10	5.26
Intermediate	85	4.25	5.00	2.51	>10	3.60	1.47	3.53	3.81	4.27	2.50	4.20	5.02	5.44	6.23	2.94	8.56	9.60	9.88	4.68
Moderate	78.7	3.42	4.37	2.30	>10	3.08	1.24	2.85	3.02	3.32	2.05	3.18	3.98	4.28	4.73	2.47	7.18	7.85	8.02	4.07
SIP	56	2.11	2.24	1.59	6.43	2.15	0.80	1.67	1.72	1.82	1.35	2.15	2.33	2.41	2.55	1.71	4.44	4.54	4.72	2.75
Eastern low sulfur																				
Stringent or intermediate	83.9	4.00	4.93	2.49	>10	3.50	1.46	3.39	3.63	4.06	2.41	4.01	4.79	5.16	5.89	2.84	8.26	9.21	9.48	4.56
Moderate	75	3.13	3.75	2.20	>10	2.92	1.18	2.57	2.71	2.95	1.92	3.18	3.59	3.79	4.17	2.33	6.57	7.11	7.26	3.87
Subbituminous																				
Stringent or Intermediate	81.2	3.70	4.62	2.47	>10	3.41	1.40	3.30	3.53	3.94	2.38	3.93	4.71	5.00	5.69	2.80	8.08	8.98	9.24	4.51
Moderate	75	3.13	3.75	2.20	>10	2.92	1.18	2.57	2.71	2.95	1.92	3.18	3.59	3.79	4.17	2.33	6.57	7.11	7.26	3.87

*Best system refers to design/operating conditions recommended in this report as appropriate to achieve close to minimum Ca/S ratios and attain high levels of SO_2 control. It is not meant to imply that one commercial system is better than another.

Note: Gas residence time for overbed feed systems was estimated by dividing bed depth by superficial velocity; no allowance was made for SO_2 formed at the top of the bed which may escape before reacting with CaO in the bed. The Mustad unit is not listed in this table because the gas phase residence time is unknown.

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between the feed sorbent size, and the actual sorbent particle size in the bed, is not rigorously known; it is possible that, although the feed sorbent size typically quoted by vendors is 1,000 to 1,500 μm mass mean, the actual size in the bed may not be that much larger than the 500 μm surface mean selected for the "best system" conditions.

In addition to Greer and Grove limestones (for the Foster-Wheeler boilers), and Tymochtee dolomite (for the FluiDyne boiler), Western 90 percent CaL (high reactivity), Bussen limestone (medium reactivity), and Menlo limestone (low reactivity) were used to estimate Ca/S requirements for systems specified by B&W (U.S.), Combustion Engineering, Johnston Boiler, B&W, Ltd. (England), and for "best system" conditions. Stones such as Grove and Menlo are included only to shown that sorbents with extremely low reactivity characteristics should not be considered due to the large quantity needed to achieve a reasonable level of SO_2 control. Western 90 percent CaL, Bussen Quarry and Menlo Quarry limestones are the sorbents used by Westinghouse in their independent assessment of industrial FBC boiler cost indicating a high, medium, and low reactivity limestone.[45] Inspection of Table 22 reveals the savings in limestone use which can be attained if "commercially-offered" design/operating conditions are modified to correspond with those recommended for "best systems." The sorbent quantities for "best system" conditions are calculated from the Westinghouse model. Each "commercially-offered" design is discussed individually in the following subsections relative to the modifications which would be necessary to operate at recommended "best system" conditions. In some cases, the substitution of recommended design/operating conditions would require redesign of the boiler to maintain capacity and/or to prevent increased particle elutriation. It is understood that the recommended conditions are different than those considered by many manufacturers, because the goal of FBC development has been to maximize system throughput. However, it is believed that modification to "best system" design/operating conditions in the future will prove cost effective. The cost impacts are discussed in Subsection 4.3.4.

3.2.1.4.1 Foster-Wheeler (Georgetown Design)--[46]The differences between this design and the envisioned "best system" conditions considered in this report are in the bed depth of 1.36 m (4.5 ft), gas velocity of 2.44 m/sec (8 ft/sec), resultant gas residence time (0.56 sec), and limestone particle size (average ≥1,000 μm). As discussed earlier and in Section 7.0, overbed feed with primary recycle is capable of efficient SO_2 control and, therefore, cannot be ruled out as the best method of SO_2 control. The most significant difference is probably the average inbed particle size of 1,000 μm (or greater) as opposed to the recommended best condition of 500 μm. If particle size were reduced and gas inphase residence time were increased slightly from 0.56 to 0.67 sec (by increasing bed depth to 1.65 m (5.4 ft) or decreasing superficial velocity to 2.06 m/sec (6.75 ft/sec)), a significant reduction in sorbent requirements could be achieved based on projections employing the Westinghouse model; as shown in column 3 of Table 22 where Ca/S ratios are cut in half by going to best system conditions using Greer limestone. Increasing bed depth would require a concomitant increase in freeboard and slightly greater capital cost, the magnitude of which would depend on boiler capacity. Reducing the superficial velocity would cause boiler derating unless the combustor cross section were enlarged, so either alternative could add to system capital cost. In this particular system, the greatest benefit could be achieved by reducing sorbent particle size. This could be done by purchasing the same or another limestone with a different particle size distribution, or the material could be crushed and/or sized onsite.

The projections of sorbent requirements for the Georgetown design using Grove limestone are high enough to eliminate use of such a low reactivity sorbent in this system. If it were to be used, particle size reduction would be recommended, in addition to increasing gas phase residence time. It is important to note that these high projections are completely independent of the abovebed feed used in the Georgetown boiler; the Ca/S ratios depend on sorbent type, particle size, and gas residence time (calculated from the expanded bed depth and superficial velocity). The overwhelming factor is the low sorbent reactivity.

3.2.1.4.2 Foster-Wheeler (Rivesville Design)--[47]This unit utilizes inbed feed with primary recycle to a carbon burnup cell. Superficial velocity ranges between 2.1 to 3.7 m/sec (7 to 12 ft/sec) with an expanded bed depth of 1.2 m (4 ft), resulting in a gas phase residence time between 0.3 to 0.57 sec. (Testing indicates that gas velocities as low as 1.1 m/sec (3.5 ft/sec) are adequate.) The Ca/S ratios shown in Table 22 are similar to but slightly higher than those noted for the Foster-Wheeler Georgetown design, which can be attributed to the lower gas phase residence time. Reduction of superficial gas velocity would enhance SO_2 removal at the expense of added boiler capital cost. Particle size reduction would also be of benefit since the sorbent in use is double-screened with a minimum size of 1,000 μm (16 mesh). Average inbed particle size may be in the range of 1,200 to 1,500 μm.

3.2.1.4.3 FluiDyne 40 in. × 64 in. Test Unit--[48]The FluiDyne unit uses inbed coal and sorbent feed with a fairly deep bed of 1.1 m (3.6 ft) and rather low superficial velocity of 0.6 to 1.8 m/sec (2 to 6.0 ft/sec) accounting for gas phase residence times greater than 0.6 sec. FluiDyne is anticipating using dolomite as a sorbent. Table 22 shows projections of sorbent needs based on Westinghouse TGA data for Tymochtee dolomite (a highly reactive sorbent), and Bussen limestone (a medium reactivity sorbent). Dolomite Ca/S molar feed ratios are characteristically lower than limestone requirements for similar operating conditions due to dolomite's higher reactivity (generally attributed to its different pore structure resulting from its magnesium content). As a result, the Ca/S ratios noted for Tymochtee dolomite are low, even lower than those listed for "best system" conditions using Western limestone at an average inbed particle size of 500 μm. However, the calcium carbonate content of Tymochtee dolomite is 60 percent or less so that total sorbent loadings would be equivalent to the case of Western limestone. Although a high reactivity dolomite may be available on a site-specific basis, the general discussion in this report emphasizes limestone use since most testing has been performed with limestone and it has wider availability. Although this tract has been taken, Tymochtee dolomite would certainly qualify as an appropriate sorbent, because of its high reactivity.

Projections of sorbent requirements were made for Bussen limestone at 1,000 μm. The resulting values are 25 to 50 percent higher than those noted for "best system" conditions using Bussen limestone. To reduce sorbent needs using limestone, particle size reduction would be effective, since other conditions are in conformance with best system conditions.

3.2.1.4.4 Babcock and Wilcox (U.S.)--[49]This design uses inbed solids feed with recycle and a gas phase residence time of 0.5 sec. The limestone top size is 9,150 μm (3/8 in.) so that average inbed particle size is probably in the range of 1,500 μm. Table 22 shows the performance which could be expected with this unit at the conditions noted, using Western, Bussen, and Menlo limestones. "Best system" performance is also listed. Sorbent requirements for the vendor specified conditions are roughly 30 to 100 percent greater than required for the "best system" conditions, regardless of limestone type or control level. The most important parameter in this case is the inbed sorbent particle size, which is larger than the recommended value of 500 μm.

3.2.1.4.5 Combustion Engineering[50] or Johnston Boiler[51]--These two units are discussed together because the specified gas phase residence times and operating temperatures are the same. Sorbent use projections for each unit are based on an inbed particle size average of 1,000 μm.* Thus, sorbent needs are the same. Both units use primary recycle although the CE unit is underbed feed and the Johnston unit is abovebed feed. (See previous discussion and FluiDyne results in Section 7.0.) To modify these two systems to best conditions, gas phase residence time would have to be increased from 0.43 sec and inbed sorbent particle size would have to be reduced. Increasing gas residence time could require some boiler redesign in both instances.

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*The actual inbed mass mean particle diameter for the CE/Great Lakes unit may be about 800 μm. This is not much different from the recommended best system condition since a surface average of 500 μm is roughly equal to a mass mean of between 600 to 700 μm.

3.2.1.4.6 B&W, Ltd.--[52]The design conditions for this unit reflect the shortest gas residence time cited by any vendor. Gas velocity is fairly high at 2.4 m/sec (8 ft/sec) with a relatively shallow bed of 0.8 to 0.9 m (2.6 to 3 ft), accounting for a gas residence time of 0.35 sec. As a result, the Ca/S ratios shown for the medium (Bussen) and low (Menlo) reactivity limestones are unacceptable, only the Western limestone indicates performance characteristics in a reasonable range (although 90 percent SO_2 reduction is projected to require a Ca/S ratio of 5.7 using the high reactivity sorbent). The modifications cited earlier for the other vendor systems would be required to attain "best system" operating conditions for SO_2 control.

3.2.1.4.7 O. Mustad and Sons--[53]Although this system appears to be run under conditions which are quite different from "best system" conditions, Mustad still predicts good SO_2 reduction at a low Ca/S ratio (75 percent at 1.5). The system has overbed feed, no recycle and an apparently low gas residence time, as well as a relatively high sorbent particle size. According to Mustad's projections, a system with these design/operating variables can meet our "best system" projections, however, further study and demonstration is required to fully assess the impact of these operating variables. Virtually no comparable data are available which have been generated under these conditions.

3.2.1.5 Other Impacts--

3.2.1.5.1 Applicability/Reliability--Industrial-sized FBC boilers are as yet unproven in extended commercial operation because fluidized-bed combustion is an emerging technology. The commercial-scale coal-fired AFBC units which are in operation (e.g., Renfrew, Johnston Boiler Company) are not being operated in typical commercial "around-the-clock" service. The AFBC units that will be used in typical service (e.g., Mustad/Enköping, B&W, Ltd. unit at the Central

Ohio Psychiatric Hospital, the Foster-Wheeler unit at Georgetown University, the crude oil heater at EXXON, the Combustion Engineering/Great Lakes unit) are not yet in operation. Such extended operation in typical service is required in order to prove AFBC reliability and to demonstrate industrial AFBC cost, energy and environmental impact. Therefore, at the present time the impacts of AFBC in comparison to conventional boilers may be somewhat underestimated or overestimated. As further information becomes available more definitive conclusions can be drawn about AFBC and its impacts.

3.2.1.5.2 Cost--The analysis of "best system" costs indicates that AFBC with SO_2 control is generally more costly than an uncontrolled conventional boiler of equal capacity by as much as 20 to 30 percent. This increment varies considerably depending on boiler capacity, coal type, SO_2 control level, and sorbent reactivity. In certain instances, controlled AFBC may be used at equal or less cost than uncontrolled conventional systems. This was found to be the case for the 8.8 MW_t unit burning low sulfur coal at any SO_2 control level, or high sulfur coal at an SIP SO_2 control level. It was also found for the 58.6 MW_t AFBC burning subbituminous coal, and is due to the equal or higher cost of pulverized coal technology at this capacity.

Another conclusion is that use of "best system" conditions can reduce the cost of FBC compared to "commercially-offered" design/operating conditions. This is due mainly to reduced operating costs due to lower limestone purchase and preparation cost and spent solids disposal costs. Adaptation of these conditions may require minor boiler redesign in some instances.

The cost trade-offs associated with decreasing total sorbent requirements by increasing gas phase residence time, decreasing sorbent particle size, or by other methods must be considered to determine the most cost-effective boiler system. For example, gas residence time can be increased by using deeper beds or lower superficial gas velocities. If deeper beds are employed, larger capacity fans and more power will be required to fluidize the bed as a result of increased pressure loss through the bed. Lowering superficial gas velocity (while maintaining constant excess air) would require beds of greater cross sectional area to maintain boiler capacity. Much more data is required to conduct a sophisticated optimization study.

Although sorbent reactivity and utilization will increase as sorbent particle size is reduced, sorbent elutriation may become severe at very fine sizes (below 500 μm) unless gas velocity is reduced correspondingly. At some point, sorbent requirements could increase unless sorbent effectiveness could be maintained by increasing primary collection efficiency and recycling large quantities of fines.

The cost of sorbent crushing and sizing must also be considered. Onsite crushing and sizing could add 15 to 40 percent to the raw limestone cost due to rejection of off-size material. However, if for example, the required Ca/S ratios are reduced from 6.0 to 3.5, a potential overall cost savings of about $\$0.90/10^6$ Btu could result (see cost sensitivity analysis in Section 4.0).

Sorbent reactivity will have a major effect on the operating cost on a site-specific basis. If a highly reactive sorbent is available in close proximity to the AFBC facility this could mean substantial cost benefit. However, if (as will likely be the case) the boiler site is not in close proximity with a highly reactivity sorbent, trade-offs must be made between the high Ca/S ratio necessary using a nearby limestone of low reactivity, or a higher reactivity limestone with a greater transportation cost. Currently, there is no surcharge for purchasing high reactivity limestones other than the incremental cost of shipment if the only available supply is remote. For an individual industry, it may be more cost-effective to use sorbent of low or average reactivity rather than pay freight costs for hauling limestone of higher reactivity from long distances.

The cost analysis in this report also indicates that the level of SO_2 control (in the range of 75 to 90 percent) does not have a large impact on FBC system cost when Eastern high sulfur coal is burned. The effect of SO_2 control level is insignificant when low sulfur coals are burned.

3.2.1.5.3 Energy Impact--The level of SO_2 control in AFBC has a minor effect on the energy impact of the total system. This is illustrated in Table 23 which shows the differential changes in boiler efficiency as FBC design/operating parameters are varied through the full range considered in this report.

TABLE 23. DIFFERENTIAL CHANGES IN BOILER EFFICIENCY VERSUS RANGE OF FBC DESIGN/OPERATING PARAMETERS

FBC design/operating parameter and range	Differential change in boiler efficiency
Sorbent reactivity – low to high*	1.83
Coal sulfur content – 0.6 to 3.5†	2.17
Boiler capacity – 8.8 to 58.6 MW_t†	1.47
SO_2 control level – moderate to stringent‡	0.58

*Stringent control, Eastern high sulfur coal.

†Stringent control, average sorbent reactivity.

‡Eastern high sulfur coal, average sorbent reactivity.

With Eastern high sulfur coal, boiler efficiency decreases by about 0.6 percent when control level is increased from moderate to stringent. This is the minimum differential change of the parameters considered. The coal sulfur content proved to have the most significant effect on boiler efficiency.

If "best system" design/operating conditions (see Subsection 3.2.1.2) for SO_2 control were implemented, this could have a favorable impact on combustion efficiency, by allowing longer residence time for carbon combustion and by recirculating char for combustion.

It is important to note AFBC energy impact relative to that of uncontrolled conventional boilers. The comparison of AFBC and uncontrolled conventional boilers showed that for any of the three smaller boilers (8.8, 22, and 44 MW_t), AFBC boiler efficiency was 1 to 3 percent higher than conventional boiler efficiency considering all optional control levels and coal types. For the larger boiler (58.6 MW_t), AFBC boiler efficiency was 1 to 3 percent lower than the conventional pulverized coal unit.

3.2.1.5.4 Environmental--In fluidized-bed combustion, the most prominent environmental impact is solid waste disposal. The "best system" design for FBC is based on minimizing the Ca/S ratio, and thus the amount of sorbent and solid waste which is necessary to achieve a given level of SO_2 reduction. Therefore, as "commercially-offered" design/operating conditions approach "best system" conditions, the environmental impact will be reduced. The amount of solid waste which is produced by a system of specific capacity is directly related to the Ca/S ratio used to achieve the necessary level of SO_2 control. The range of solid waste produced by systems discussed in this report is 123 kg/hr (270 lb/hr) to 3,873 kg/hr (8,533 lb/hr), representing the 8.8 MW_t boiler using low sulfur coal achieving a moderate control level and the 58.6 MW_t boiler using high sulfur coal achieving stringent control, respectively.

The data presented previously in Table 22 illustrate that sorbent requirements can vary significantly depending on system design/operating conditions and sorbent reactivity. Considering a sorbent of reasonable reactivity, Ca/S requirements can be reduced significantly if "best system" design/operating conditions are substituted for "commercially-offered" conditions. For instance, if Greer limestone is considered, the Ca/S ratio can be reduced to 3 or slightly less using "best system" design/operating conditions as opposed to values between 4 and 5 for "commercially-offered" conditions and stringent or intermediate SO_2 reduction (based on projections from the Westinghouse model). If the Ca/S ratio is reduced from 5 to 3, spent solids waste quantities will fall by approximately 30 percent.

The environmental concerns associated with the disposal of the waste are due to the leachate which is generated and the heat release properties of the waste upon initial contact with water. The pH of the leachate is high, and the total dissolved solids content is above drinking water standards. Calcium and sulfate are also present in the leachate at concentrations above drinking water standards.[54]

These facts do not present an insurmountable problem, but do suggest that appropriate care must be taken in disposing of the residue. It is not expected at this time that trace elements will typically be present in the leachate at levels greater than 10 times the drinking water standards, the level at which the residue would be considered "hazardous" (toxic) under the Resource Conservation and Recovery Act (RCRA). This conclusion, however, must be confirmed with further testing.

Air emissions are also affected by applying "best system" conditions. The SO_2 control system in FBC affects NO_x emission reduction and add-on particulate control devices. Some evidence indicates that NO_x emissions are lower over a partially sulfated bed than over an inert bed.[55,56] To this extent, the SO_2 removal system may enhance NO_x reduction. Generally, particulate control is compatible with the SO_2 removal system. However, finer particles of high resistivity (sorbent derived) will be elutriated as sorbent particle size is reduced to minimize sorbent feed requirements. It is not anticipated that this will impact the ability of final control devices in meeting the optional particulate control levels considered in this report.

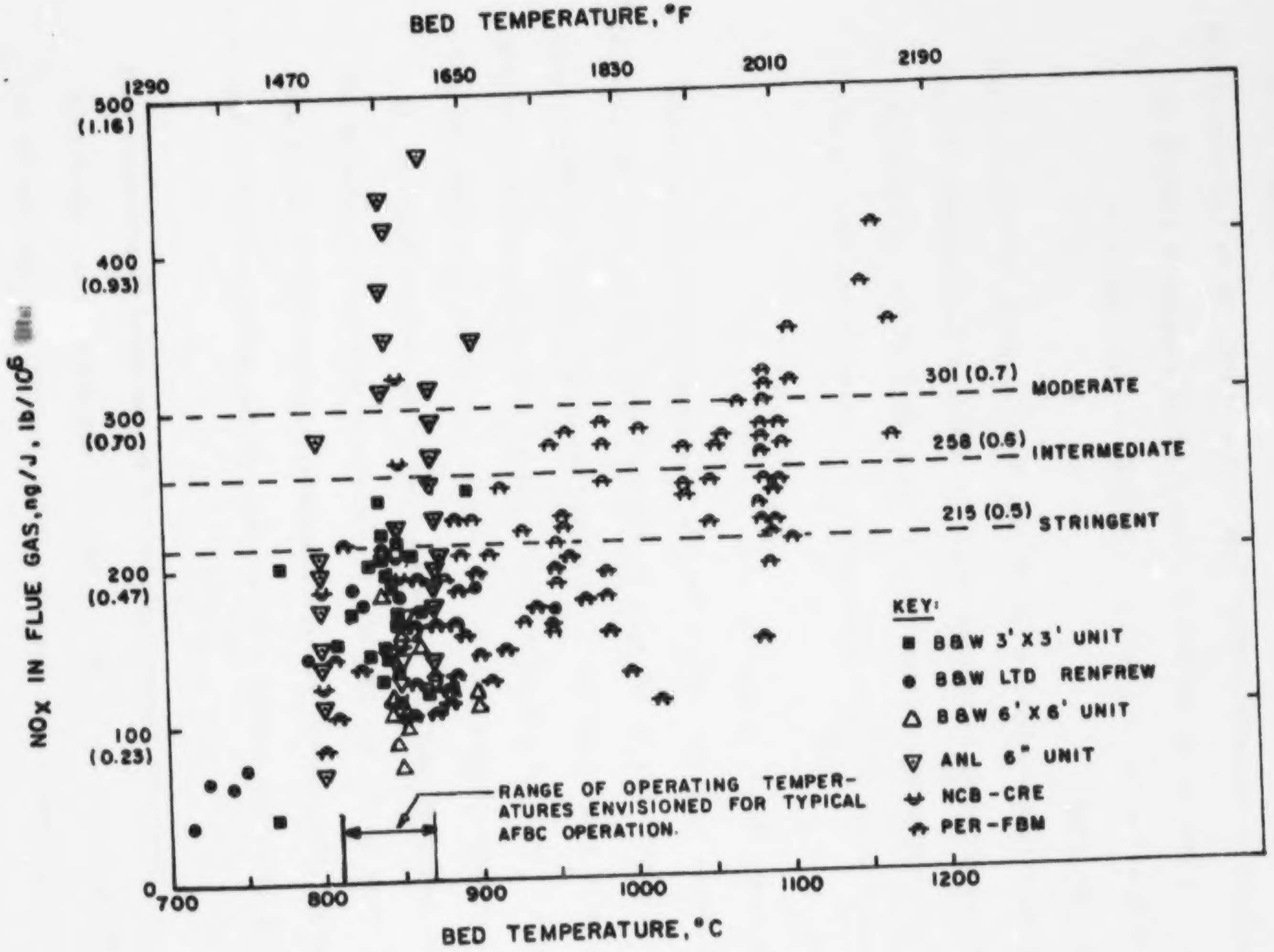
3.2.2 NO_x Emissions

Based on existing experimental FBC NO_x emission data, the "best system" of NO_x control requires no special modifications from "best system" design/operating conditions for SO_2 control. An AFBC designed for effective SO_2 control should be capable of simultaneously achieving the optional levels of NO_x control.

3.2.2.1 Moderate Reduction Controls--

The moderate level of control for NO_x emissions to be supported using fluidized-bed combustion is 301 ng/J (0.7 $lb/10^6$ Btu). This level has typically been met in most runs in virtually all experimental FBC units (including units as small as 0.15 m (6 in.) diameter) under normal operating conditions burning coal (bed temperatures less than 1000°C, excess air levels from 10 to 100 percent, stable operation, and gas residence times of 0.2 sec or longer). In larger AFBC units (3 MW_t and larger), NO_x emissions have rarely exceeded 301 ng/J (0.7 $lb/10^6$ Btu), except at high temperatures (above 1100°C) which are representative of carbon burnup cell temperatures but not of typical industrial FBC operation.

Figure 27 illustrates the predominance of NO_x emission measurements that fall below 301 ng/J (0.7 $lb/10^6$ Btu). Some ANL measurements are above this level at operating temperatures less than 900°C (1650°F) but the results are not representative because the AFBC unit was small (6 in. in diameter) and an



*THESE POINTS ARE ESTIMATED FROM DATA REPORTED IN ppm, THUS THE ACCURACY OF THESE POINTS IS ASSUMED TO BE ± 30%

Figure 27. Summary of NO_x data from experimental AFBC units.

inert bed was used.[57] Several other measurements from Pope, Evans, and Robbins are above the moderate level but operating temperatures were greater than 1100°C (2010°F), a value characteristic of CBC operation.

Table 24 summarizes the range of NO_x emission values reported by several investigators, along with key operating conditions in existence during the testing. It is noted that gas residence times were generally below 0.67 sec, which should be appropriate for effective SO_2 and NO_x control. Excess air rates are generally around 20 percent which is considered the nominal rate for current and future AFBC designs. The range in operating conditions noted (temperature, gas residence time, and excess air) encompasses the design/operating conditions previously tabulated for "commercially-offered" systems in Table 21 (see Subsection 3.2.1.4). In general, commercially-offered designs are planned to operate at bed temperatures between 800° to 900°C (1472° to 1652°F), will use gas residence times between 0.4 to 0.5 sec, and will operate with excess air rates between 15 to 25 percent. Possible exceptions are units being developed by FluiDyne and O. Mustad and Sons. FluiDyne may use gas residence times up to 2.0 sec, and bed temperatures as low as 700°C (1292°F) although these may just be experimental extremes. Mustad is building systems with staged combustion. Either system should be capable of effective NO_x control, possibly better than the other systems noted.

Comparing the experimental conditions with the "commercially-offered" conditions, it is apparent that commercially-offered systems should be capable of controlling NO_x to levels within those shown experimentally. If gas residence times are increased to correspond with that noted for "best systems" (0.67 sec), then improved NO_x control should be possible. Regardless, the moderate NO_x level should be achievable without design or operating modifications.

TABLE 24. SUMMARY OF EXPERIMENTAL NO_x DATA FROM ATMOSPHERIC FBC TEST UNITS*

Investigator	Unit size	Range of operating conditions				Range of NO_x emissions observed ng/J ($lb/10^6$ Btu)	Comments
		Temperature °C (°F)	Excess air %	Gas phase residence time (sec)	Fuel nitrogen %		
B&W, Ltd. Renfrew, Scotland	10 × 10 ft ~12 MW_t (40×10^6 Btu/hr)	690 - 900 (1274 - 1652)		~0.3 - 0.7	1.1	70 - 198 (0.17 - 0.46)	This is one of the largest FBC units for which NO_x data exists. The reported data (approximately 11 tests for NO_x) are all below the stringent control level of 215 ng/J (0.5 $lb/10^6$ Btu).
B&W Alliance, Ohio	6 × 6 ft ~7 MW_t (24×10^6 Btu/hr)	835 - 899 (1535 - 1650)	9.9 - 44.4	0.30 - 0.57	1.03 - 1.34	77 - 185 (0.18 - 0.43)	This range of NO_x emissions was reported for 56 individual tests (see Table in Section 7), each of 100 to 1,000 hours duration. Most testing was performed with excess air rates between 16 to 20 percent.
B&W Alliance, Ohio	3 × 3 ft ~1.9 MW_t (6.5×10^6 Btu/hr)	770 - 894 (1418 - 1642)	nominal excess O_2 = 3%	0.13 - 0.21	0.76 - 1.23	47 - 262 (0.11 - 0.61)	The maximum of 262 ng/J was noted once out of 30 tests. The next highest reading was 236 ng/J (0.55 $lb/10^6$ Btu) so that 29 of 30 tests met the intermediate level of NO_x control. 20 of 30 tests met the optional stringent level, even though gas residence times were generally below 0.2 seconds.
Pope, Evans, and Robbins	1.5 × 6 ft 3.2 MW_t (11×10^6 Btu/hr)	804 - 1021 (1480 - 1870)	5 - 25	0.13 - 0.29		87 - 228 (0.20 - 0.53)	64 of 65 reported NO_x test results fell below the optional stringent control level although gas residence time was low, generally about 0.20 seconds.
		1021-1176 (1870-2147)		0.13-0.29		190-405 (0.44-0.94)	The experimental temperature range is significantly above that envisioned for typical AFBC operation. Nevertheless, 75 percent of the recorded data are below 301 ng/J (0.7 $lb/10^6$ Btu).
National Coal Board	3 × 1.5 ft ~1.3 MW_t (4.5×10^6 Btu/hr)	749 - 849 (1380 - 1560)	≤29	0.26 - 1.76	1.3 - 1.5	120 - 323 (0.28 - 0.75)	The maximum level was noted to drop to 191 ng/J (0.44 $lb/10^6$ Btu) during the same individual test run. The average emission based on 17 reported values calculates to 215 ng/J (0.5 $lb/10^6$ Btu). 9 of 17 recorded NO_x values were below the optional stringent control level. The maximum gas residence time of 1.76 sec is atypical; most were in the range of 0.5 seconds.
Argonne National Laboratory	6 in. diameter bench scale ~0.3 MW_t	718 - 900 (1325 - 1650)	6 - 25	0.22 - 1.0	1.11 - 1.31	70 - 435 (0.16 - 1.01)	Although this unit is a small bench scale test unit, over 2/3 of reported NO_x data (115 individual tests) were below the optional moderate level of NO_x control.

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*Based on NO_x emission data shown in Section 7.

NO_x control at this level should be routine and should not contribute any additional cost, energy, or environmental impact above that associated with normal AFBC boiler operation.

3.2.2.2 Stringent Reduction Controls--

"Best systems" should require no special design or operation beyond that for "best system" SO_2 control. However, this needs to be confirmed in future experimentation and actual commercial operation.

The stringent level of control targeted for FBC is 215 ng/J (0.5 $lb/10^6$ Btu). A review of existing emissions data indicates some individual small pilot-scale experimental systems have been able to meet these requirements without any deliberate efforts to control NO_x (see Figure 27 and Table 24). For instance, PER has reported NO_x emissions ranging between 86 to 172 ng/J (0.2 to 0.4 $lb/10^6$ Btu) during operation of their FBC and FBM test units.[58] The design of these units is similar to that expected in first generation industrial FBC boilers although gas residence times were shorter than used in current designs.* Testing of the B&W 3 ft × 3 ft unit has consistently demonstrated NO_x emissions less than or equal to 236 ng/J (0.55 $lb/10^6$ Btu) and a minimum emission of 47 ng/J (0.11 $lb/10^6$ Btu).[59] This minimal value was measured at a gas residence time of 0.62 sec, the longest reported during this test series. In general, the stringent level of NO_x control has been met in over half of the runs on smaller facilities.

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*As shown in Figure 27, PER has conducted extensive experimentation in the FBM unit at temperatures higher than envisioned for typical AFBC operation, and as a result, NO_x emissions higher than the optional stringent level of 215 ng/J (0.5 $lb/10^6$ Btu) have been recorded.

The stringent level has been met consistently on the larger AFBC units which have been operated to date (Renfrew, B&W 6 ft × 6 ft unit).* The effect of AFBC boiler capacity on NO_x emission rate is illustrated in Figure 28. The full range of NO_x test results is included in the vertical bar shown for each test unit. Not only do emissions decrease as the size of the facility increases, but also, the two larger units had no reported NO_x values above the stringent level of 215 ng/J (0.5 lb/10^6 Btu). These two units operate at typical conditions seen for commercial systems (see Table 20), and indicate that the stringent level should be achieved without system modifications, or added cost, energy, or environmental impact.

Increasing gas residence times to 0.67 sec (from the average value of about 0.5 sec noted for these two larger units) could result in even lower NO_x emissions.

There are probably technical and economic upper limits to extending gas residence time since deeper beds would be required. In addition, incremental reductions in NO_x emission rate might diminish as residence time increases. Excess air rates between 10 and 20 percent, as normally cited for FBC operation, are probably the minimal or best levels for NO_x control. Operating at lower excess air levels might reduce combustion efficiency.

Some experimentation has been performed to assess the benefit of applying NO_x combustion modification techniques to FBC. Research at ANL showed NO_x emissions between 43 to 129 ng/J (0.1 to 0.3 lb/10^6 Btu) when combustion air was fed in stages to FBC.[60] Although early results support the capability of two-stage combustion in lowering NO_x emissions, combustion modification should

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*Preliminary results from Rivesville (30 MW_e unit by Foster-Wheeler and PER) indicate NO_x emissions as low as 86 ng/J (0.2 lb/10^6 Btu).

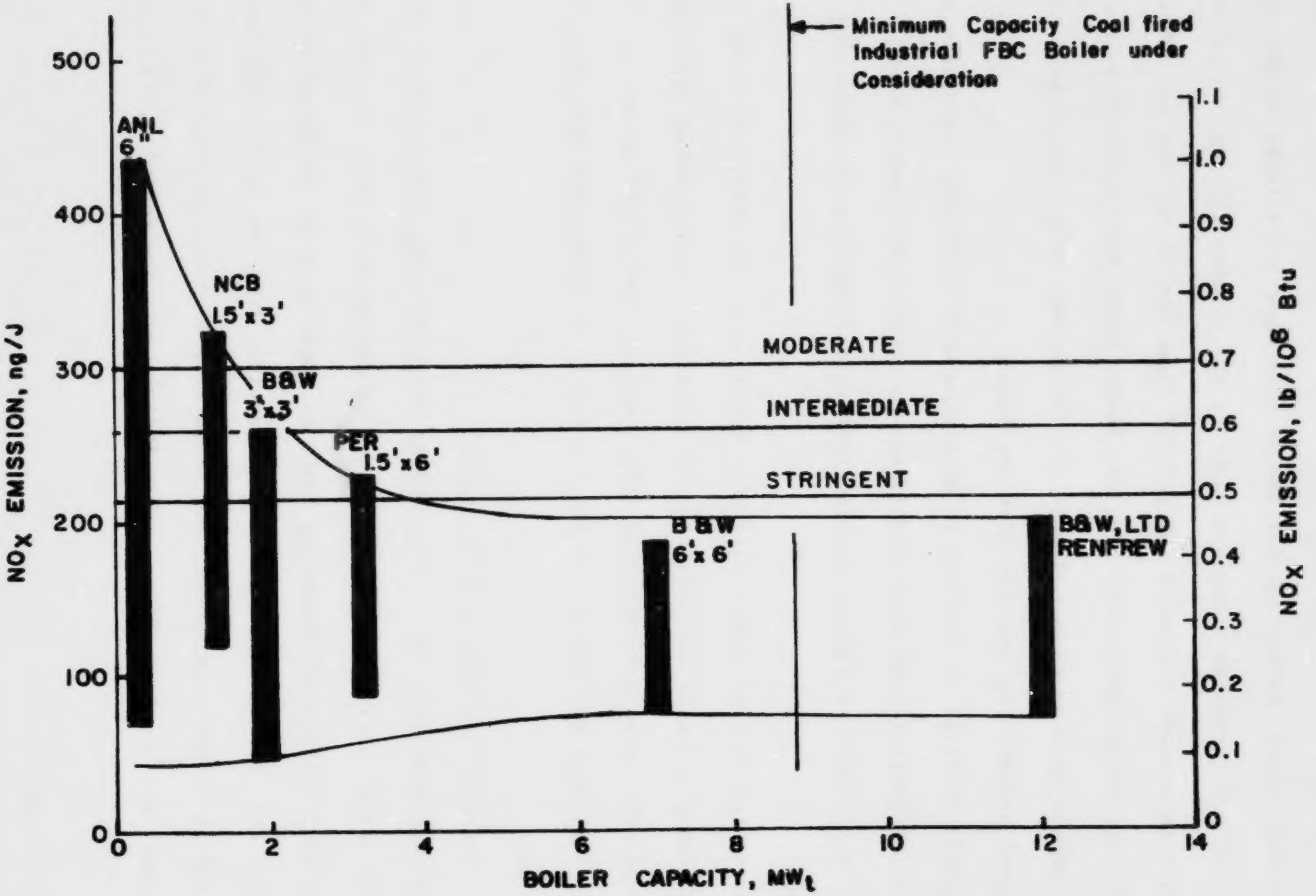

Figure 28. NO_x emissions from experimental FBC units as a function of capacity.

not be necessary to meet 215 ng/J (0.5 lb/10^6 Btu). Also, it is not considered an available control technology for FBC at this time. The level of SO_2 reduction in FBC may establish a minimum limit to the primary air rate in two-stage combustion to prevent excessive CaS and H_2S formation in the bed and subsequent SO_2 formation in the freeboard. The SO_2 formed above the bed would be ineffectively removed because of the absence of high concentrations of sorbent and the minimal sorbent SO_2 reaction time available.

Further analysis is required to determine whether staged combustion and flue gas recirculation or other modifications could significantly improve NO_x control in FBC boilers without causing operational problems or increasing other emissions.

The reliability of controlling NO_x emissions at the stringent level on a 24-hour average basis during long-term operation is not certain, since data from large AFBC units are currently very limited. The actual mechanisms which strongly influence NO_x control in FBC are not fully identified and understood at this time.

3.2.2.3 Intermediate Reduction Controls--

The intermediate level of NO_x control which is being considered is 258 ng/J (0.6 lb/10^6 Btu). A large percentage of NO_x emission data recorded at all existing AFBC test units (including units as small as 6 in. in diameter) have been below this level.* As discussed above, data from the larger AFBC facilities (operating at normal primary cell bed temperatures) have been consistently below the intermediate level of 258 ng/J (0.6 lb/10^6 Btu). In addition, during testing of the somewhat smaller 36 in. × 18 in. CRE unit by the British National Coal Board, all but four of the measured NO_x values were below this level.[61] Based on existing data, it is expected that industrial FBC boilers will be capable of supporting an intermediate NO_x control level without incorporation of special design/operating features.

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*The PER FBM data above this level were recorded in experimentation conducted at bed temperatures much higher than envisioned for typical AFBC operation.

3.2.3 Particulate Emissions

Necessary particle control efficiencies to meet the optional control levels under consideration are shown in Table 19, Subsection 3.1.4. Uncontrolled emissions refer to the loading downstream of the FBC primary cyclone, which is considered an integral part of the FBC system. The ranges in particle loading and mass median diameter at the outlet of the primary cyclone are also shown in Table 19.

It is essential to note that final particulate control technology has not been demonstrated in FBC to date. In the near future, testing is planned at EPA's Sampling and Analysis Test Rig, Georgetown University, and Rivesville, West Virginia. There are some data available for primary cyclone inlet and outlet loadings (as shown in Sections 7.0 and 2.0), but it is important to expand the data base.

3.2.3.1 Moderate Reduction Controls--

The moderate particulate control level to be supported using fluidized-bed combustion and add-on controls is 107.5 ng/J (0.25 lb/10^6 Btu). Emission control techniques which could be used to reduce particulate emissions to this level include multitube cyclones (MC), electrostatic precipitators (ESP), and fabric filters (FF). A comparison of these controls is presented in Table 25 illustrating relative differences in cost, energy impact, environmental impact, reliability, applicability, and other factors, by boiler capacity. Wet scrubbers

TABLE 25. APPLICABILITY OF FINAL PARTICULATE CONTROL DEVICES TO ACHIEVE MODERATE CONTROL AT 107.5 ng/J (0.25 lb/10^6 Btu) FOR COAL-FIRED FBC INDUSTRIAL BOILERS

Boiler capacity MW_t (10^6 Btu/hr)	Final control device	Technological ability to meet control level	Cost	Applicability* in meeting control level	Energy impact	Environ-mental impact	Boiler operation or safety	Reliability	Status of development with respect to controlling FBC emissions	Multi-pollutant control capability	Adaptability to new FBC boilers	Compatibility with FBC	Overall ranking
58.6 (200)	MC	B	B	A	B	A	A	B	C	A	A	A	A
	FF	B	D	D	B	A	B	B	C	A	A	B	C
	ESP	B	D	D	A	A	A	B	D	B	A	C	C
	WS	B		D									D
44 (150)	MC	B	B	A	B	A	A	B	C	A	A	A	A
	FF	B	D	D	B	A	B	B	C	A	A	B	C
	ESP	B	D	D	A	A	A	B	D	B	A	C	C
	WS	B		D									D
22 (75)	MC	B	B	A	B	A	A	B	C	A	A	A	A
	FF	B	D	D	B	A	B	B	C	A	A	B	C
	ESP	B	D	D	A	A	A	B	D	B	A	C	C
	WS	B		D									D
8.8 (30)	MC	B	A	A	B	A	A	B	C	A	A	A	A
	FF	B	D	D	B	A	B	B	C	A	A	B	C
	ESP	B	E	D	A	A	A	B	D	B	A	C	C
	WS	B		D									D

*For moderate control, ESP's or FF's would be inapplicable because they represent overdesign.

Notes: Rating System - Each control device is rated by a letter code (A = best; B = good; C = acceptable; D = poor; E = inappropriate) relating to each factor listed in the table. The overall ranking applies to all factors listed in the text.

MC - Multitube Cyclone
FF - Fabric Filter
ESP - Electrostatic Precipitator
WS - Wet Scrubber

are itemized, but are not considered as an appropriate option for particulate control in FBC. Therefore, not all of the items have been rated for wet scrubbers.

Considering the tenfold range of emissions downstream of the FBC primary cyclone (215 to 2,150 ng/J) and resulting overlap in efficiency requirements to meet stringent, intermediate, and moderate levels, the comparison given in Table 25 is for efficiency requirements between 50 and 80 percent. If greater than 80 percent efficiency is required to meet a moderate level of 107.5 ng/J (0.25 $lb/10^6$ Btu), then the comparison in Table 25 does not apply. The discussion of intermediate and stringent levels indicates the trade-offs associated with using different particulate removal devices at control efficiencies greater than 80 percent.

A rating system from A to E is assigned to compare control devices capable of meeting a moderate standard, as explained in the footnotes to Table 25. The overall ranking indicates that the best system for moderate control is the multitube cyclone. In general, fabric filters and ESPs are inappropriate because they represent overdesign and unnecessary cost for moderate particulate reduction. The relative cost of add-on control devices is shown in Figure 29 based on the analysis in Section 4.0. ESP costs for SIP control were estimated in Section 4.0 to be significantly higher than multitube cyclone cost for moderate control. This fact, and the results shown in the figure indicate that a multitube cyclone is the low cost device.

Several of the categories are interrelated, such as technological ability, reliability, and compatibility with FBC. Since final control devices have not been demonstrated on FBC units, none of these factors can be explicitly defined.

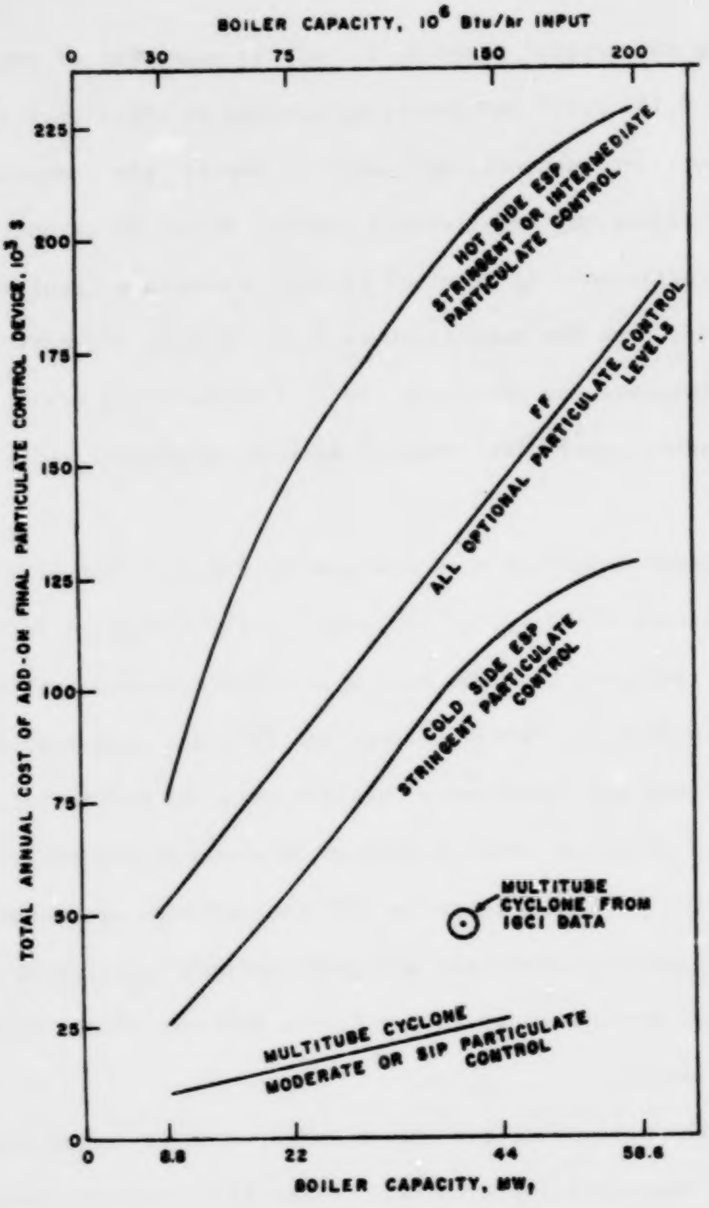

Figure 29. Cost of final particulate control for AFBC industrial boilers.

All devices should have the technical capability to meet the moderate level so they have each been assigned a B rating. An A rating was not assigned because demonstration of these devices on FBC boilers has not occurred. There could be problems with fine particulate removal in multitube cyclones, blinding or bag fires in fabric filters, and unsuitably high particle resistivity for ESP use. Therefore, compatibility with FBC is questionable, mainly for ESPs or fabric filters. Reliability must be proven for all systems in extended testing. Therefore, all devices were assigned a B rating in this category.

The energy impact of fabric filters or multitube cyclones is slightly greater than ESPs because of higher pressure drop. The environmental impact of any of the three systems should be similar because an equivalent amount of solid material is removed at a specific efficiency and material is handled in dry form.

No major problems with boiler operation and safety are foreseen, other than with fabric filter use where the potential for bag fires must be assessed. Also, since fabric filters do not have natural bypass capabilities, inadequate fabric cleaning procedures could result in sudden pressure drop increases that might affect the operation of the boiler.

Considering multipollutant control capability, use of any add-on final particle control device should not have any detrimental effect on SO_2 or NO_x control capability in FBC. ESPs were assigned a B rating in this category because FBC particle resistivity data indicate that ESPs must be operated as hot-side installations for suitable performance. Consequently, there may be otherwise condensable trace elements which would escape a hot-side ESP. However, the associated environmental impact should be negligible.

Adaptability of add-on final particulate devices to new FBC boilers should not be a general problem for any specific device. Therefore, all systems have been rated equivalently. Adaptability will be most significantly influenced by site-specific conditions.

3.2.3.2 Stringent Reduction Controls--

The stringent control level for particulate reduction is 12.9 ng/J (0.03 lb/10^6 Btu). Based on particulate emissions ranging from 215 to 2,150 ng/J (0.5 to 5 lb/10^6 Btu) with mass mean size of 5 to 20 μm after the primary cyclone, the final collection efficiency requirements range between 94 to 99.4 percent. The most applicable devices for control at this level are fabric filters and ESPs.

Multitube cyclones are not capable of routinely achieving this level of control, and wet scrubbers have not received serious consideration because of the generation and handling of liquid wastes. In addition, wet scrubbers would have to operate a high pressure drops to attain high efficiency particle collection.

Although fabric filters and ESPs should be capable of stringent particulate control, there are uncertainties which preclude a clear cut selection of either device as the best system for application to FBC boilers due to the early stage of development. These factors have been mentioned in the previous subsection, but they deserve reemphasis here. Primarily, final control device performance on FBC boilers has not been demonstrated to date. This assessment is based upon the performance of these devices on conventional system particulate emissions. Their performance on FBC should not be grossly different from that on conventional boilers burning low sulfur coal. However, in the case of ESPs, particle resistivity may cause performance problems. PER and TVA measurements shown in Section 2.0 indicate that hot-side installation is required for ESP use. ESP reliability may be poor depending on variability in coal. Much more experimentation is necessary to confirm that hot-side ESPs would function well. In the case of fabric filters, there is a potential for bag blinding due to lime hydration or bag fires. The influence of factors such as caking, bag cleaning, and bag durability have not been explored. Until these uncertainties and possible problems are confirmed or refuted in actual testing, a clear-cut decision between the two devices is not possible.

There are some specific advantages or disadvantages that could influence the choice of a fabric filter or ESP. Primarily, fabric filters are a lower cost system than hot-side ESPs (see Figure 29), based on costs quoted for conventional boilers burning low sulfur coal. The total annual cost of the fabric filter is 15 to 30 percent less than the hot-side ESP. When the total FBC system costs are added, the cost difference becomes insignificant, because, at worst, add-on device cost approaches 10 percent of total boiler system cost. This is shown in detail in Section 4.0.

ESPs should have slightly lower energy impact due to negligible pressure drop. However, as efficiency requirements become more stringent, the advantage disappears. An ESP may be preferred from the standpoint of boiler operation and safety since sudden back pressure increases with improperly cleaned fabric filters could cause operating problems.

Neither fabric filters or ESPs have significant multipollutant control capability, but fabric filters would have an advantage over hot-side ESPs because they would capture condensable trace elements and organics in the range of 100° to 150°C which would pass through a hot-side ESP uncontrolled.

Fabric filters may be more adaptable than ESPs to small capacity boilers because of lower capital cost and less operational variability and complexity. Operating a hot-side ESP to overcome resistivity problems requires handling significantly larger gas volumes than would be necessary with use of a fabric filter. Coal and sorbent type could vary appreciably, especially at smaller boiler installations, resulting in differences in particle resistivity which would affect ESP collection efficiency. Assuming that hot-side ESP operation is essential, fabric filters should be more compatible with small capacity FBC boilers.

All of the important factors influencing the choice of the best system of particulate control at the stringent level are summarized in Table 26. Complete ratings are provided only for ESPs and fabric filters, since these devices alone are considered technically capable of stringent control. The remaining factors of concern are environmental impact and adaptability to new FBC boilers. There should be no significant difference in ESP or fabric filter use for either of these considerations.

3.2.3.3 Intermediate Reduction Levels--

The intermediate standard under consideration for particulate removal is 43 ng/J (0.1 $lb/10^6$ Btu). The required final efficiency to meet this level ranges between 80 to 98 percent. Best system selection in the range of 94 to 98 percent follows the discussion presented for stringent control. In the range of 80 to 94 percent, fabric filters, ESPs, or multitube cyclones could be applicable depending on site-specific conditions.

System comparisons and applicability are similar to the previous discussions for moderate and stringent control, depending on the proximity of required control efficiency to 80 percent or 94 percent, respectively. Multitube cyclones

TABLE 26. APPLICABILITY OF FINAL PARTICULATE CONTROL DEVICES TO ACHIEVE STRINGENT CONTROL AT 12.9 ng/J (0.03 lb/10^6 Btu) FOR COAL-FIRED INDUSTRIAL BOILERS

Boiler capacity MW (10^6 Btu/hr)	Final control device	Technological ability to meet control level	Cost	Applicability in meeting control level	Energy impact	Environmental impact	Boiler operation or safety	Reliability	Status of development with respect to controlling FBC emissions	Multi-pollutant control capability	Adaptability to new FBC boilers	Compatibility with FBC	Overall ranking
58.6 (200)	FF	A	A	A	B	A	B	C	D	A	A	B	B
	ESP	B	B	A	A	A	A	C	D	B	A	C	B
	MC	E	E	E									E
	WS	D	E	E									E
44 (150)	FF	A	A	A	B	A	B	C	D	A	A	B	B
	ESP	B	B	A	A	A	A	C	D	B	A	C	B
	MC	E	E	E									E
	WS	D	E	E									E
22 (75)	FF	A	A	A	B	A	B	C	D	A	A	B	B
	ESP	B	C	A	A	A	A	D	D	B	A	D	B
	MC	E	E	E									E
	WS	D	E	E									E
8.8 (30)	FF	A	A	A	B	A	B	C	D	A	A	B	B
	ESP	B	D	A	A	A	A	D	D	B	A	D	B
	MC	E	E	E									E
	WS	D	E	E									E

Notes: Rating System - Each control device is rated by a letter code (A = best; B = good; C = acceptable; D = poor; E = inappropriate) relating to each factor listed in the table. The overall ranking applies to all factors listed and discussed in the text.

FF - Fabric Filter
ESP - Electrostatic Precipitator
MC - Multitube Cyclone
WS - Wet Scrubber

might be applicable for the low end of this range if mass median particle size is greater than 10 μm. Under this condition, multitube cyclones would be the low cost device. Otherwise, fabric filters would be the low cost alternative (see Figure 29). Again, it is important to consider the uncertainties due to the lack of demonstration on FBC boilers.

3.3 OTHER FUELS

Data on emissions from fluidized-bed combustion of residual and distillate oil or natural gas are limited. Therefore, it is premature to discuss the rationale or ability to support optional standards for oil or gas combustion in FBC. Also, the extent of oil or natural gas use in FBC is uncertain, but is not expected to be widespread.

The summary (Section 3.4) presents emission reduction requirements necessary for SO_2, NO_x, and particulate, under the three optional standards. Requirements for SO_2 control are listed for residual and distillate oil and NO_x emission reduction requirements are shown for coal and oil together. It is projected that fluidized-bed combustion of oil should be capable at least of meeting the optional standards for SO_2 and NO_x applicable for coal combustion. It is possible that more stringent NO_x levels could be achieved due to lower fuel oil nitrogen content. SO_2 and NO_x emissions from combustion of natural gas are expected to be low, due to low sulfur and nitrogen content of natural gas, and low combustion temperature.

3.4 SUMMARY

The candidate best systems of emission reduction associated with FBC are summarized in Tables 27 through 29 for SO_2, NO_x, and particulate emissions.

TABLE 27. OPTIONAL SO_2 CONTROL LEVELS AND REQUIRED EFFICIENCIES

Fuel and boiler capacity MW_t (10^6 Btu/hr)	Sulfur content (%)	Uncontrolled SO_2 emission ng/J (lb/10^6 Btu)	Level of emission control and efficiency required to achieve that level ng/J (lb/10^6 Btu)			Best system of SO_2 control - Ca/S ratio requirements								
			Stringent	Intermediate	Moderate	Stringent Control			Intermediate Control			Moderate Control		
			90% removal or control efficiency required to achieve 86 ng/J (0.2 lb/10^6 Btu)	85% removal or control efficiency required to achieve 86 ng/J (0.2 lb/10^6 Btu)	75% removal or control efficiency required to achieve 516 ng/J (0.2 lb/10^6 Btu)	Sorbent reactivity								
						High	Average	Low	High	Average	Low	High	Average	Low
Coal 8.8 to 58.6 (30 to 200)	3.5	2425 (5.64)	90	85	78.7	2.3	3.3	4.2	2.1	2.9	3.8	1.8	2.5	3.4
	0.9	533 (1.24)	83.9	83.9	75	2.0	2.8	3.7	2.0	2.8	3.7	1.6	2.2	3.2
	0.6	512 (1.19)	83.2	83.2	75	2.0	2.7	3.6	2.0	2.7	3.6	1.6	2.2	3.2
Residual oil 44 (150)	3.0	1350 (3.14)	90	85	75		3.3*			2.9*			2.5*	
Distillate oil 4.4 (15)	0.5	219 (0.51)	60.8	60.8	60.8		1.2*			1.2*			1.2*	

*Estimated - not based on actual data from oil-fired units.

TABLE 28. OPTIONAL NO_x CONTROL LEVELS

Fuel and boiler capacity MW (10^6 Btu/hr)	Uncontrolled NO_x emission ng/J (lb/10^6 Btu)	Level of emission control and NO_x reduction efficiency required to achieve that level ng/J (lb/10^6 Btu)			Control device required
		Stringent 215 (0.5)	Intermediate 258 (0.6)	Moderate 301 (0.7)	
Coal and oil					
4.4 - 58.6 (15 - 200)	430* (1.0)	50	40	30	AFBC†

*Highest reported value for FBC using calcium-based sorbent.

†Ability of AFBC to achieve the stringent level of control without some adjustment of design/operating conditions (to excess air values as low as 15%, and to gas residence times as high as 0.67 sec) must be confirmed by further data on large AFBC units.

TABLE 29. OPTIONAL PARTICULATE CONTROL LEVELS AND REQUIRED EFFICIENCIES (AFTER PRIMARY CYCLONE)

Fuel and boiler capacity MW (10^6 Btu/hr)	Uncontrolled particulate emission ng/J (lb/10^6 Btu)	Particle size average MMD (μm)	Level of emission control and efficiency required to achieve that level ng/J (lb/10^6 Btu)			Control device recommended*		
			Stringent 12.9 (0.03)	Intermediate 43 (0.10)	Moderate 107.5 (0.25)	Stringent	Intermediate	Moderate
Coal								
8.8 - 58.6 (30 - 200)	215 - 215.0 (0.5 - 5.0)	5 - 20	94 - 99.4	80 - 98	50 - 95	ESP or FF	ESP, FF or MC	MC

*Selection of device will depend upon efficiency requirements, particle size, boiler capacity, and tradeoffs in the economic and energy requirements of each device. (See Tables 3-5 and 3-6.)

FF - Fabric Filter
ESP - Electrostatic Precipitator
MC - Multitube Cyclone

3.4.1 SO_2

The best SO_2 control system in AFBC is the one which minimizes sorbent requirements, energy impact, and cost impact, and simultaneously maintains the control level of concern. Based on review of experimental results, estimates of Ca/S ratio requirements for best SO_2 control are given in the last columns of Table 27, for SO_2 removal efficiencies ranging between 75 to 90 percent. The values selected are average values calculated from several experiments which were conducted using average sorbent particle sizes close to 500 μm and gas phase residence times close to 0.67 sec. The average Ca/S ratio from the experimental results shown in Table 27 is considered representative because SO_2 reduction results were reported for sorbents of low and high reactivity. The Ca/S ratios shown are used in the remainder of this report to assess cost, energy, and environmental impact. These values were chosen instead of model projections for specific sorbents (i.e., Western 90 percent CaL, Bussen, and Menlo) because the experimental Ca/S ratios are taken from a wide data base and should be more representative of the sorbent requirements of a typical user. Also, the Menlo sorbent reactivity is probably too low for practical use.

As SO_2 removal requirements become more stringent, air pollution impact will be minimized, but the impact of disposing of large volumes of sulfated bed material will increase. However, the spent stone is in dry form, which should simplify handling.

Reliability of performing within the optional SO_2 standards has been proven in a wide variety of pilot-scale FBC boilers. The most critical factors are selection of a suitable sorbent, use of appropriately small particle sizes, and operation with sufficiently long gas phase residence times. Sorbent characteristics have been studied thoroughly and are documented in a number of references. The FBC SO_2 control model developed by Westinghouse illustrates the dependence of Ca/S molar feed ratios on FBC design and operating conditions.

3.4.2 NO_x

Experimentation has illustrated the potential of FBC to support any of the three optional levels. The major concern is that additional data from large AFBC units are necessary to confirm the ability of AFBC to reliably achieve the stringent level of control. Data from large units are currently limited, but the data which do exist (B&W 6 ft × 6 ft, Renfrew) support the ability of AFBC to meet the stringent level.

3.4.3 Particulate

Particulate reduction under all three control options should be possible in FBC systems by using suitably designed and operated conventional add-on particulate control devices. This has not yet been demonstrated, because suitably large AFBC units with final particle control have not been operated for sufficiently long periods. However, control of particulates from AFBC should be similar to control in conventional boilers burning low sulfur coal.

The most important factors in selecting a device are cost and reliability. For stringent or intermediate control, fabric filters are the low cost device (unless mass median particle size is large enough to allow the use of multitube cyclones for lower efficiency requirements under intermediate control). For moderate control, multitube cyclones are the low cost device.

When total system cost is considered (i.e., the AFBC boiler with all auxiliaries plus final particulate control) cost differences as a function of the final particulate control device employed are small because the cost of the add-on device is at most 5 to 10 percent of the total annual boiler cost.

Reliability of final particulate control for FBC must be proven in large-scale testing. Existing data indicate that ESPs will have to be operated as hot-side installations because of high particle resisitivity. ESP performance could be impacted by variability in coal and sorbent characteristics, a factor which could be especially important in smaller capacity boilers. Fabric filter performance and reliability is also uncertain due to potential problems with bag blinding, and bag fires.

These uncertainties must be explored in full-scale testing. In the near future, testing is planned at the EPA's Sampling and Analysis Test Rig, Rivesville, and Georgetown University.

Since one of the implicit purposes of FBC is to avoid liquid waste production, use of wet scrubbers has not been given serious consideration.

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4.0 COST IMPACT OF IMPLEMENTING BEST SYSTEMS OF EMISSION CONTROL

4.1 INTRODUCTION

4.1.1 Background

Industrial-sized FBC boilers are as yet unproven in extended commercial operation because fluidized-bed combustion is an emerging technology. The commercial scale coal-fired AFBC units which are in operation (e.g., Renfrew, Johnston Boiler Company) are not being operated in typical commercial "around-the-clock" service. The AFBC units that will be used in typical service (e.g., Mustad/Enkoping, the B&W Ltd. unit at Columbus State Hospital in Ohio, the Foster Wheeler unit at Georgetown University, the crude oil heater at Exxon, the Combustion Engineering/Great Lakes unit) are not yet in operation. Such extended operation in typical service is required in order to prove AFBC reliability and to demonstrate industrial AFBC costs. Therefore, at the present time the cost of AFBC in comparison to conventional boilers may be under- or overestimated. As cost data from first generation commercially-operated FBC boilers become available, more accurate cost estimates can be developed. Second generation systems may be more cost-effective because of design and operating improvements.

Most of the discussion in this section centers on SO_2 control to assess the influence of meeting optional levels of desulfurization on FBC cost. NO_x control is considered intrinsic to the system and no specific costs are readily identified. Particulate control will be attained with add-on devices similar to those used for conventional combustion with low sulfur coal; hence, cost of particulate control should not be very sensitive to FBC design variations - although operating data are needed to substantiate this assumption.

Costs are presented for "best system" designs and potential savings compared to "commercially-offered" AFBC designs are estimated (see Section 3.2.1.5). Since "commercially-offered" boiler designs do not typically incorporate the conditions (0.67 second gas residence time, 500 μm average in-bed sorbent particle size) that are felt to represent the "best" SO_2 control system, and since data for SO_2 control efficiency at these conditions are limited, the need for confirmation of the "best" SO_2 control system costs by large scale AFBC operation is especially important.

The cost values presented in this section are budget estimates for a new technology operating under hypothetical conditions and are probably accurate to within ±30 percent. Even wider variation could exist depending on site specific conditions, and, therefore, these results are presented only as an indication of the benefits or penalties of using FBC in place of conventional technology on a broad basis. The results are not intended to provide a basis for selecting one technology over another for a specific industrial application; they are meant to reflect trends which are valid only for a preliminary comparison of two different technologies. Therefore, the thrust of the analysis is not the generation of absolute cost values, but a comparison of the cost of FBC with SO_2 control under various operating conditions against cost of conventional boilers without SO_2 control.

The sensitivity analysis presented later takes a prominent role in the overall discussion as an analysis of the effect on cost of several possible operational modes. Again, it is recommended that the absolute costs presented later be treated cautiously.

A more plausible comparison could be made if the conventional boilers also included some form of SO_2 control (e.g., flue gas desulfurization). Because this report is one part of an overall system analysis of pollution control for industrial boilers being done by EPA, all pollution control options (FBC was one of eight options) were separately compared with uncontrolled conventional reference boilers. The results of this study are interesting in that even without considering the added cost of SO_2 control for conventional systems, there may be some cost advantage to FBC over conventional boilers in certain size ranges or if low sulfur coals are burned.

4.1.2 Data Sources

The FBC cost estimates developed in this section are based on vendor quotes and are mid-1978 dollars. These vendor quotes were supplemented by reference to cost data developed by PEDCo for conventionally-fired boilers of the same capacity.[1] Other recent cost estimates for industrial fluidized-bed combustion were reviewed and two estimates, one by Exxon Research and Engineering,[2] the other by A.G. McKee,[3] are included for comparative purposes. In addition, the results of an independent AFBC industrial boiler cost assessment prepared by Westinghouse Research and Development under EPA sponsorship are also included.[4] The Westinghouse costs were partly derived using information supplied by GCA, but in-house Westinghouse FBC cost data were used to determine total capital, operating, and annual costs. Westinghouse did not solicit vendor quotes for boiler cost.

4.1.3 Data Uncertainties

The cost variation among these estimates is at least partially a function of the wide variation in atmospheric FBC designs among different vendors. Certain differences which are important include:

- methods of coal and limestone handling and feeding;
- freeboard height;
- bed depth;
- heat transfer tube placement and orientation;
- use of fly ash recirculation or carbon burnup cell (use of the latter will probably be very limited in industrial boilers);
- coal and limestone particle size; and
- normal fluidization rates.

Several methods of coal and limestone feeding are being advocated and these methods require further investigation to determine which feed technologies will provide adequate dispersion at minimal cost. Load variation (turndown) is another area where several techniques are being developed and the feasibility of these must also be studied.

There is some debate relative to maintenance requirements in fluidized-bed boilers. Equipment of particular concern includes in-bed boiler tubes and coal feeders. Boiler tubes in the bed may be items of high maintenance due to the possibility of fluctuating oxidation/reduction zones near coal feed points. However, maintenance of in-bed tubes may be reduced due to the relatively low and constant temperatures in the bed compared to conventional boilers and corrosion/erosion may be reduced by suitable design (e.g., by not placing tubes in the immediate vicinity of coal injection points). In-bed coal feeders may be a high maintenance item due to potential clogging and erosion; overbed coal feeders can avoid the plugging and erosion problems, but could necessitate double screening of the coal to avoid increased emissions.

Because the impact on maintenance cost of these system components cannot be assessed from the current data base, neither a penalty or advantage can be assigned to FBC maintenance requirements relative to conventional systems. As operating experience is gained from industrial scale plants, detailed estimates of maintenance costs can be developed. For this analysis it is assumed that maintenance requirements and boiler life expectancy are similar for FBC and conventional systems.

The costs presented here assume that all three levels of NO_x control can be achieved with no impact upon FBC cost. This assumption is based upon the fact that most NO_x data from all experimental AFBC units are below the intermediate level of 258 ng/J (0.6 lb/10^6 Btu).* Data from the larger (>250 kg/hr coal) AFBC units are consistently below 215 ng/J (0.5 lb/10^6 Btu) at typical primary cell bed temperatures.* Therefore, the stringent level can be expected to be reliably attained with little or no adjustment to standard design and operating conditions. In practice, any such adjustments may have some impact on capital and operating costs, but it is not possible to quantify at this time. Finally, it is not expected that SO_2 control variations will have a significant influence on NO_x control capability or cost. In practice, the increased gas residence times desired for good SO_2 control should tend to reduce NO_x emissions.

4.1.4 Major Contributors to Emission Control Costs for SO_2

4.1.4.1 General Comments--

An AFBC industrial boiler is an integrated energy production/SO_2 control technology. Consequently, certain equipment items and operating costs can not be discretely isolated as part of the steam raising system on the SO_2 control system.

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*See Sections 2 and 3.

The forced draft fan and associated fan power fall into this category. Also included are the primary cyclone (for recycle of unreacted fine sorbent and unburned carbon), the induced draft fan and associated power use, and ancillary equipment for feeding and discharge of bed material. The subsequent discussion of contributors to the capital and operating cost of SO_2 control should be considered in the context of these factors.

4.1.4.2 Capital Costs Related with SO_2 Removal--

The major capital costs associated with use of AFBC as an emission control technique are the boiler, the limestone handling and feeding system, and spent solids handling and disposal. Ancillary equipment items normally required for AFBC and conventional systems and of similar cost are coal handling, induced draft fan, water treatment equipment, instrumentation, stack, etc. Common items which are of higher cost in FBC systems are the coal feeders and the forced draft fan.

The boiler cost will depend on several design variables. Influential factors are: shop versus field erection, freeboard height, bed configuration, heat transfer design, carbon recirculation design, and load following technique. Several designs are available which incorporate different combinations of these variables. At this stage of development, no single design is expected to dominate the market.

Limestone handling capital cost depends on sorbent storage and feed rate requirements, which in turn depend on SO_2 control level, sorbent reactivity, coal sulfur content, and boiler size. The amount of limestone storage at a specific site will depend upon available delivery frequency and possibly haulage rates. Limestone feeding capital cost will depend on design (i.e., separate or combined with coal feeding), and will vary primarily as a function of boiler capacity.

Spent solids handling capital cost is a direct function of limestone feed rate. The most significant cost contributors are onsite storage and disposal site capital cost. The latter will vary depending on the disposal site requirements.

Coal feeders represent a possible capital cost increase for AFBC as compared to conventional firing. This cost differential results from the critical need for even fuel distribution throughout the cross-section of a fluidized-bed combustor. The major design classifications are overbed screw feeders, overbed spreader stokers, and underbed pneumatic injectors.

The forced draft fan in AFBC handles slightly lower air volumes than the conventional coal-fired boilers under consideration (due to lower excess air) but must overcome about 3 times the pressure drop encountered in a conventional coal-fired boiler.* Most of the additional pressure drop in the AFBC is across the grid plate and the bed.

4.1.4.3 Operating Costs Related with SO_2 Removal--

Limestone purchase and solid waste disposal costs are the most important direct operating cost variations associated with supporting optional SO_2 control levels with FBC boilers. These costs can be reduced by reducing sorbent feed requirements through careful boiler design and operation. The power required to run the forced-draft fan is also a contributor.

Although limestone reactivity has a potentially important impact on sorbent feed requirements for a given SO_2 control level, an industrial FBC user may not always have the flexibility to choose a highly reactive sorbent because it may be located at such a distance that haulage costs are excessive. Each individual FBC user will have to balance the tradeoff between purchasing high grade limestone versus operating with lower reactivity stones at higher sorbent feed rates. (The resulting sorbent feed and spent solids rates will have a moderate impact on the capital cost of materials storage.) This means that the cost of best SO_2 control can vary from site to site. Generally, an AFBC user at a typical site should have available to him at least one source of sorbent of reasonable reactivity. The cost estimates presented in this chapter consider a range of sorbent feed rates, based upon a range of reasonable sorbent reactivities. If, indeed at a given site, the only quarries within an economically transportable distance have extremely nonreactive sorbents, then AFBC might not be the SO_2 control option of choice for that particular site.

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*See analysis of energy requirements in Section 5.0, Subsection 5.2.3.

Site specific factors also influence the operating cost of spent solids disposal. The most important are disposal site location, and applicable waste disposal regulations.

Research is currently being performed to determine methods to: (1) minimize solid waste from FBC boilers; (2) identify and abate the potential environmental impact; or (3) find suitable byproduct uses. FBC residue characteristics which are of most concern are leachate pH, Ca^{++}, $SO_4^{=}$, total dissolved solids and heat release during hydration.[5,6] If FBC spent solids require special handling/disposal (e.g., fixation at the plant, or imperviously-lined containment), handling costs could increase significantly. These factors might influence plant siting or could add to the cost of an in-city plant that pays to have its wastes hauled to disposal.

Special handling/disposal problems are not anticipated under the Resource Conservation and Recovery Act.[7] Trace elements are not typically present in the leachate at levels greater than 10 times the drinking water standards. This is the level at which the residue would be considered "hazardous" (toxic) under the Resource and Conservation and Recovery Act. Leachate concentrations must be confirmed through further testing on waste from commercial size units.

Byproduct uses for FBC solid waste are being investigated by L. John Minnick,[8] the U.S. Department of Agriculture,[9] Westinghouse,[10] Ralph Stone and Co., Inc.,[11] and TVA,[12] and several universities and private concerns. Cost or siting limitations might be reduced if the waste can be utilized.

Electricity is required for operation of coal and limestone handling and feeding, spent solid withdrawal and cooling, FD and ID fans, and boiler water circulation and treatment. The FD fan is the major user, and consumes about half of the total auxiliary power requirement.* Operation of the primary particulate recycle device (normally a cyclone) will require minimal fan energy because the pressure drop is low (<15 cm (6 in.) w.g.).

4.1.5 Cost Related with Final Particulate Removal

The cost of particulate control for FBC boilers is significant but should be similar to particulate control on conventional boilers burning low sulfur coal. Uncontrolled emissions (downstream of the primary cyclone) are similar to conventional systems.‡ Flue gas volumes are slightly less for FBC boilers in comparison to conventional boilers of the same capacity because excess air rates are lower and efficiencies are somewhat higher. Moderate control (50 to 80 percent reduction) may be achieved with use of multitube cyclones.† Stringent control (94 to 99.4 percent reduction) requires installation of a fabric filter or ESP.[*] Intermediate control (80 to 94 percent reduction) will require any one of these three devices depending on actual efficiency necessary and other site specific conditions.[*]

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*See Section 5.0.

†See Section 3.0.

‡See Section 2.0 and 7.0.

Particulate removal cost will be influenced by SO_2 control because of limestone elutriation. Particulate control needs may increase with sorbent addition, but incremental loadings are uncertain, so that the significance of cost variation is questionable.

4.1.6 Most Important Cost Items

A summary of important capital and operating cost items associated with FBC boiler operation and emission control is shown in Table 30. The most significant cost impact which varies as a function of SO_2 control level is the direct operating cost of limestone purchase and solid waste disposal. Total FBC system cost will also be influenced by particulate control requirements.

TABLE 30. MAJOR COST CONTRIBUTORS TO FBC BOILER CAPITAL AND OPERATING COST

Capital	FBC boiler (replaces conventional boiler)
	Forced draft and induced draft fan
	Coal feeding
	Primary and final particulate collection
	Limestone storage and handling
	Spent solids storage, handling, and disposal
Operation	Coal purchase
	Limestone purchase
	Spent solids disposal
	Forced draft fan power
	Final particulate collection

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[*] See Section 3.0.

The analysis of "best system" costs indicates that AFBC with SO_2 control is generally more costly than an uncontrolled conventional boiler of equal capacity by as much as 30 percent. This increment varies considerably depending on boiler capacity, coal type, SO_2 control level, and sorbent reactivity. In certain instances, controlled AFBC may be used at equal or less cost than uncontrolled conventional systems. This was found to be the case for the 8.8 MW_t unit burning low sulfur coal at any SO_2 control level, or high sulfur coal at an SIP SO_2 control level. It was also found for the 58.6 MW_t AFBC burning subbituminous coal, and is due to the equal or higher cost of pulverized coal technology at this capacity.

Another conclusion is that "best system" designs can reduce the cost of FBC compared to "commercially-offered" design/operating conditions (see Section 4.3.4). This is mainly due to reduced operating costs. Capital costs may be higher or lower depending on the alterations necessary and the specific design of interest.

The analysis also indicates that the level of SO_2 control (in the range of 75 to 90 percent) does not have a large impact on FBC system cost when Eastern high sulfur coal is burned. The effect of SO_2 control level is insignificant when low sulfur coals are burned. A more important consideration in determining the cost impact of SO_2 control is sorbent reactivity. This results because sorbent quantities vary through a greater range as a function of the extremes of sorbent reactivity considered in this study.

4.2 GROUNDRULES FOR DEFINING COST BASIS

The AFBC costs presented are for a grass roots boiler installation in the midwest. The facility battery limits are from, but not including, the coal receiving equipment to, and including, the stack and onsite spent solids storage. The cost of land for offsite spent solids/ash disposal is included in the annualized disposal cost. The water treatment facility is included but piping for the steam to and from the process area is not.

4.2.1 Capital Costs

New facilities have been costed in conformance with guidelines presented by PEDCo.[13] Direct costs include all equipment, installation, and land. Indirect costs include engineering costs, construction and field expenses, contractor's fees, startup, performance testing, contingencies, and working capital. Indirect costs are estimated as a percentage of direct costs with the factors used for FBC estimates summarized in Table 31.

TABLE 31. VALUES SELECTED FOR ESTIMATING INDIRECT FBC CAPITAL COSTS FOR NEW FACILITIES

Cost item	Value selected
Engineering	10% of installed costs
Construction and field expenses	10% of installed costs
Contractor's fee	10% of installed costs
Startup	2% of installed costs
Contingencies	20% of total direct and indirect costs
Working capital	25% of the total annual operation and maintenance costs

4.2.2 Operating and Annualized Costs

The annual cost of owning and operating an FBC industrial boiler consists of operation and maintenance, overhead, and capital charges. Operation and maintenance covers all costs incurred to operate the FBC system on a daily basis, and includes utilities, raw materials, operating labor, routine maintenance and repairs, fuel purchase, and spent solids disposal cost. Table 32 summarizes the unit cost values used to estimate FBC operation and maintenance costs.

TABLE 32. UNIT COST VALUES USED TO ESTIMATE ANNUAL OPERATION AND MAINTENANCE COSTS FOR FBC INDUSTRIAL BOILERS

Cost factors	Unit cost*
Direct labor, \$/man-hour	12.02†
Supervision, \$/man-hour	15.63‡
Maintenance labor, \$/man-hour	14.63†
Electricity, mills/kwh	25.8§
Untreated water, \$/1,000 gal	0.12
Process water, \$/1,000 gal	0.15
Cooling water, \$/1,000 gal	0.18
Boiler feed water, \$/1,000 gal	1.00
Coal, High S, $\$/10^6$ Btu (\$/ton)(Eastern)	0.74 (17.00)#
Low S, $\$/10^6$ Btu (\$/ton)(Eastern)	1.16 (29.00)#
Low S, $\$/10^6$ Btu (\$/ton)(Wyoming)	0.42 (6.75)#
No. 2 fuel oil, $\$/10^6$ Btu	3.00
No. 6 fuel oil, $\$/10^6$ Btu	2.21
Natural gas, $\$/10^6$ Btu	1.95
Lime, \$/ton (bulk, FOB works)	32.00**
Limestone, \$/ton (bulk, FOB quarry)	6.00**
Limestone, \$/ton (bulk, FOB plant)	8.00††
Spent solids disposal, \$/ton offsite	40.00††

*All costs are in June 1978 dollars.

†Engineering News-Record, June 29, 1978, pp 52-53, Average for Chicago, Cincinnati, Cleveland, Detroit and St. Louis.

‡Estimated at 30 percent over direct labor rate.

§EEI members publication for June 1978, Average for Boston, Chicago, Indianapolis, Houston, San Francisco, and Los Angeles.

#Coal Outlook, 7/18/78 issue, Spot market prices.

**Chemical Marketing Reporter, June 19, 1978.

††See subsection 4.3.2 for discussion of limestone purchase and spent solids disposal costs.

The cost of offsite spent solids disposal is based on a common unit cost factor recommended for use in all current technology assessment reports being done as part of EPA's industrial boiler systems study. Consequently, no credit has been allowed for possible cost savings associated with dry material handling and disposal. Possible income from sale for byproduct uses has not been considered. The unit cost value is used to determine total annual disposal cost and includes amortized capital associated with land purchase, disposal site preparation, and necessary offsite equipment. Transportation and necessary labor are also included.

Coal costs do not include transportation to be consistent with other technology assessment reports. Transportation cost was included in the limestone purchase cost since this is a cost specific to AFBC technology.

Since all of these costs can vary considerably from site to site depending on transport distance, coal and sorbent type, and waste disposal requirements; the impact of that variation is estimated in the cost sensitivity analysis in Subsection 4.3.8.

Overhead costs (payroll overhead and plant overhead) have been included and cover services such as administration, safety, engineering, legal, medical, payroll, benefits, recreation, and public relations. The values are:

Payroll overhead = 30 percent of operating labor

Plant overhead = 26 percent of labor and materials

Equipment and installation costs, expressed as annualized capital charges, are calculated by applying an appropriate capital recovery factor. To facilitate comparison with the estimates made by PEDCo for conventional boilers, an expected rate of return of 10 percent and life expectancy of 30 years were selected.

Modified and reconstructed facilities are not considered in this cost analysis. The economics of such installations are not certain and could be misleading if presented on a generalized basis. The cost of retrofit is highly dependent on site-specific conditions.

4.2.3 Specific Vendor Quotes

Several vendors were contacted to request capital and operating cost information for FBC industrial boilers. Vendors contacted included Foster-Wheeler,[14] Babcock & Wilcox,[15] Babcock & Wilcox, Ltd (England),[16] Johnston Boiler,[17] Energy Resources Company,[18] and Combustion Engineering.[19] Cost information for the four standard AFBC boilers has been received from three vendors, referred to here as Companies A, B, and C.* The information from Companies A and B was used by GCA to develop cost estimates for AFBC boiler plants according to the format recommended by PEDCo. The latter vendor quote was received late in the study and was used only as an internal check of the values presented later. Subsection 4.2.3.4 discusses the results of this comparison.

4.2.3.1 Company A — Basis of FBC Boiler Costs--

Capital and operating cost data were provided for AFBC boilers of the following capacity:

8.8 MW_t (30×10^6 Btu/hr) — full shop fabrication
2 MW_t (75×10^6 Btu/hr) — field erection of shop fabricated modules
44 MW_t (150×10^6 Btu/hr) — full field erection
58.6 MW_t (200×10^6 Btu/hr) — full field erection

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*The vendor quotations are treated anonymously due to the major additions and alterations which were necessary to adjust the costs to comply with the costing format recommended for this study. In the final analysis, the basic boiler cost is only a small part of the total annual system cost.

Company A noted that the smallest capacity boiler was below the range that they intend to build so that this cost will not be presented here. Capital costs were quoted to include the following equipment (the limits represent that equipment a boiler manufacturer would normally provide):

1. FBC cells
2. Steam generation pressure system parts
3. Flue duct dampers
4. Underbed plenum
5. Air heater and/or economizer
6. Refractory insulation and lagging
7. Structural steel
8. Platform stairways, rails, etc.
9. Ignition system
10. Valves and trim
11. Forced draft fan and motor drive
12. Induced draft fan and motor drive
13. Overbed fuel feed system*
14. Limestone injection system
15. Bed material extraction and cooling system

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*Experimentation by FluiDyne (see Section 7.0) has shown comparable high efficiency SO_2 removal for both in-bed and above-bed fuel/sorbent feeding systems (in their 18 in. × 18 in. unit) when primary recycle is practiced. This result is observed despite the fact that, in an overhead feed system, some SO_2 may be released above the bed and, thus, not have a full residence time within the sorbent bed. Therefore, the cost of the overhead type of feed system should be consistent with achieving "best system" SO_2 control, using the same Ca/S ratios that would be projected assuming that all of the SO_2 is released near the bottom of the bed. The cost sensitivity analysis in Subsection 4.3.8 indicates an added total system cost of $\$0.40/10^6$ Btu output if capital cost is underestimated by 20 percent. This should encompass the added cost of an in-bed fuel/sorbent feed system. However, it is not anticipated that in-bed feed is necessary, as long as primary recycle of elutriated sorbent/char is practiced.

16. Control and safety system
17. Mechanical collection system
18. Fly ash reinjection system
19. Steam coil air heater
20. Instrumentation

The quote does not include the following equipment:

1. Foundation
2. Motor control center
3. Instrument control panel
4. Intermediate wiring and tubing
5. Building
6. Bulk material plant receiving (coal, oil, limestone)
7. Storage bunkers, (coal, limestone, residue)
8. Auxiliary fuel storage
9. Boiler feed water treatment
10. Boiler feed water pumps
11. Spent material (residue) transfer
12. Stack
13. Intermediate piping and valves (including feed water control valve)

Representative operating conditions associated with the FBC boilers provided by Company A are:

- Steam Pressure: 100 to 1,000 psi, increasing with boiler capacity
- Fluidization Velocity: 6 to 8 ft/sec
- Approximate Expanded Bed Depth: 3 ft to 4 ft
- Approximate Gas Phase Residence Time: 0.4 to 0.67 sec
- Excess Air: 20 percent

This vendor noted that Ca/S molar feed ratio would have negligible impact on cost of equipment provided for each boiler.

4.2.3.2 Company B - Basis of FBC Boiler Costs--

Company B quoted capital equipment costs for a shop fabricated AFBC boiler of 8.8 MW (30×10^6 Btu/hr) capacity. A complete boiler unit includes:

1. FBC cell
2. Under bed plenum
3. Ignition system
4. Coal feed hoppers
5. Limestone feed hoppers
6. Coal and limestone variable speed above-bed screw feeders
7. Flue duct dampers
8. Steam trim
9. Feedwater regulator
10. Forced draft fans and drives
11. Induced draft fan and drives
12. Instrument and control panel
13. Primary particulate control equipment with reinjection
14. Stack and transition
15. Materials feed bins

Coal-fired boilers provided by Company B typically operate with a fuel to steam efficiency between 81 to 83 percent. Design steam pressure for the unit quoted is 150 psi. Steam is produced at a rate of approximately 11,350 kg/hr (25,000 lb/hr). Excess air is typically in the range of 20 percent.

This company is planning to use limestone with a particle size distribution of 85 percent >1190 μm (16 mesh) and a top size of 2380 μm (8 mesh). Expanded bed depth is approximately 0.84 m (32 to 34 in.). Gas phase residence time is about 0.45 sec based on a superficial velocity of 1.8 m/sec (6 ft/sec).

The items included in these two listings are different, reflecting the fact that Company B is providing completely shop fabricated systems. Company A's systems are larger and require partial or complete field erection so that certain items such as the stack and instrument control panel are considered as extra equipment.

4.2.3.3 Other Capital Costs--

To supplement and complete the cost estimates provided by Company A and B, the following equipment costs were based on data supplied to PEDCo for conventional boilers:

- Stack (Company A only);
- Boiler feedwater treatment and circulation equipment; and,
- Coal handling.

Costs for materials handling equipment (limestone, spent solids, and ash) were estimated based on correspondence with other vendors.[20-23]

4.2.3.4 Company C - Cost Estimates

Capital cost estimates for the two larger AFBC boilers (44 and 58.6 MW_t) were received from a third vendor, but at a late juncture in the preparation of this report. The costs were adjusted to include all necessary auxiliary equipment, direct and indirect installation, and contingencies, for consistency with the procedures used in this report. Total capital charges were between 10 and 25 percent lower than those reported in the following analysis. Capital costs were annualized and added to direct operating costs and overhead. The resultant total annual charges for vendor C were 5 to 7 percent lower than the

AFBC costs estimated based on Company A information. We elected to consider this third estimate only as a check on the data developed in the detailed cost analysis which follows.

4.2.4 Other FBC Boiler Cost Estimates

Westinghouse Research and Development is currently preparing a study entitled "Effect of SO_2 Emission Requirements on Fluidized-Bed Boilers for Industrial Applications: Preliminary Technical/Economic Assessment." The preliminary results of their cost analyses are included in Subsection 4.3.7.1.[24] Westinghouse used the cost basis defined in this study but based costs on in-house information and sources other than boiler vendor quotes.

Other reports on industrial FBC boiler costs have been reviewed for comparative purposes. These include reports by EXXON,[25] and A.G. McKee.[26] The detailed cost assumptions used in these reports are noted in Appendix B. Some adjustments were made to the estimates to attain comparability with the basic assumptions used in this report. A description of these adjustments also appears in Appendix B. These costs, as adjusted are shown in terms of $\$/10^6$ Btu in subsection 4.3.7.2. They are compared with our estimates based on quotes by Companies A and B.

4.3 COST ANALYSIS FOR IMPLEMENTING BEST SYSTEM OF SO_2 CONTROL

Derivation of the cost of AFBC purchase and operation with SO_2 control required use of a two-tiered approach. The costs which are independent of (but not necessarily divorced from) the three optional control levels (stringent, intermediate moderate) on which the study is based represent the first tier. These basic costs which are assumed to vary only with boiler capacity and coal type are presented in Appendix A, Tables A-1 through A-12. The second tier is composed of those costs which vary as a function of SO_2 control level, and sorbent reactivity, in addition to coal sulfur level and boiler capacity. The costs which are dependent on the degree of sulfur dioxide retained are presented in Appendix C, Tables C-20 through C-24. While the summation of these costs represents total cost of operation of an AFBC with SO_2 control, the first tier of costs is not intended to represent the cost of an uncontrolled AFBC boiler. This procedure was followed solely for ease of computation in estimating the cost of AFBC operation under the several options considered in this report.

The second tier of costs includes the following components:

- Capital costs
  - Limestone storage, conveying, and screening
  - Spent solids/ash conveying, and storage
- Operating costs
  - Limestone purchase
  - Spent solids/ash disposal
  - Electricity for operation of all auxiliary equipment (excluding building utilities such as lighting, heating, ventilating, and air conditioning)

4.3.1 Capital Costs

Limestone handling capital cost assumes a storage bin capacity for 14 days at full load. Double screening and pneumatic conveying equipment is also necessary. Limestone crushing is performed at the quarry. Limestone feeding capital cost was included in the basic AFBC boiler costs (Appendix A, Tables A-1 through A-12) because no significant cost variation with respect to control level or sorbent reactivity is anticipated.

The major equipment items necessary for spent solids handling are a storage bin (10 days capacity at full load) and pneumatic conveying. Capital cost of spent solids withdrawal and cooling should only vary significantly as a function of boiler capacity and coal feed rate, and is included in the boiler costs developed in Appendix A. The spent solids handling costs presented as part of the SO_2 control cost do not include capacity for particulate matter collected in the final particulate control device. (The incremental cost for elutriated fines handling is presented in the discussion of particulate control costs.) Equipment was sized for SO_2 control by assuming 90 percent of all sorbent and ash which enter the FBC combustor are removed at the spent solids withdrawal point. The particulate matter downstream of the primary cyclone then ranges between 365 and 1850 ng/J (0.85 to 4.3 lb/10^6 Btu) which is within the envelope of experimental results discussed in Sections 2.0 and 3.0 of this report.

Volumetric limestone and spent solids storage requirements were estimated using hourly processing rates derived from material balance considerations. As discussed above, a factor of 0.9 was applied to the spent solids rates to determine storage requirements for SO_2 control. Capital cost estimates were prepared based on correspondence with several equipment vendors. Storage bins account for about 80 percent of total materials handling capital cost. They include ancillary equipment such as dust control equipment, feed and exit ports, access ladders, etc. They are assumed to be fabricated of 0.64 to 0.95 cm (1/4 to 3/8 in.) carbon steel.[27] Below 283 m^3 (10,000 ft^3) capacity, units are shop fabricated and delivered to the site.[28] Above this capacity, field erection is required. For shop fabricated limestone storage bins, a variable unit cost ranging from \$353/m^3 (\$10/ft^3) down to \$282/m^3 (\$8/ft^3) was applied as storage capacity increased.[29] Above 283 m^3 (10,000 ft^3), the estimated cost for field erection is \$383/m^3 (\$11/ft^3) to provide a limestone bin, equivalent to a shop fabricated bin in stage of completion.[30] (This accounts for added field labor costs and contingencies.)

Slightly higher unit costs were used for spent solids storage to account for any incremental cost incurred due to the higher temperature of the waste material (such as: added wall thickness, wall linings, etc.). For shop fabrication, unit costs of \$392/m^3 (\$11.10/ft^3) ranging down to \$304/m^3 (\$8.60/ft^3) were used. For field erection of units above 283 m^3 (10,000 ft^3), a unit cost of \$431/m^3 (\$12.20/ft^3) was applied.

The cost of remaining capital equipment items for sorbent and spent solids handling was generally estimated in proportion to storage costs. A factor of \$4.40/kg/hr (\$2.00/lb/hr) of limestone feed capacity was added to account for screening equipment (i.e., \$6,600 for screening if estimated limestone requirements are 1500 kg/hr (3300 lb/hr)). A factor of 10 percent was added to this subtotal to account for pneumatic limestone handling equipment. A factor of 15 percent was added to spent solids storage cost to account for pneumatic spent solids handling equipment. To determine total installed capital cost of materials handling facilities, 35 percent was added for direct installation cost, 30 percent was added to total direct cost to estimate indirect installation requirements, and 20 percent was added to total installed costs for contingencies (see Table 31; engineering, construction and field expenses, contractor's fee, and contingencies).

4.3.2 Operating Costs

Annual operating costs are based on a load factor of 0.6.* Limestone purchase and spent solids disposal are based on the following unit costs:[31]

- Limestone purchase – $8.82/10^6 g ($8.00/ton) FOB plant
- Spent solids disposal – $44.10/10^6 g ($40.00/ton) offsite

The limestone purchase unit cost includes delivery to the FBC plant site. The spent solids disposal unit cost is based on a transport distance of 20 miles to the disposal site. The cost includes all necessary operating costs and the amortized capital cost of land and equipment.

Electricity required for operation of all auxiliary equipment is shown in Table C-24 as a function of boiler capacity, coal type, SO_2 control level, and sorbent reactivity. Annual electricity costs are included in the total FBC cost estimate by assuming a unit cost of 2.58¢/kWh.

The costs presented subsequently in terms of $/10^6 Btu output are based on the boiler efficiency ratings estimated in Section 5.0.

4.3.3 Cost of Best Systems of SO_2 Control

The incremental costs discussed in Subsections 4.3.1 and 4.3.2 are itemized in Appendix C in Tables C-20 (Total Turnkey Cost of Limestone Handling and Storage), C-21 (Total Turnkey Cost of Spent Solids Handling and Storage), C-22 (Annual Cost of Limestone Purchase), C-23 (Annual Cost of Spent Solids Disposal), and C-24 (Annual Cost of Electricity). The corresponding cost associated with uncontrolled conventional boilers is also shown.†

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[*] The relationship between AFBC system cost and plant load factor is shown in Figure 47 in Subsection 4.3.8.4.

[†] Throughout this chapter, the cost of uncontrolled conventional boiler systems or components is based on the results of the PEDCo cost study.[32]

Table 33 presents the total annual cost of AFBC industrial boilers using the "best system" of SO_2 control as identified in Section 3.0, (i.e., gas phase residence time of 0.67 sec, bed depth of 1.2 m (4 ft), superficial gas velocity of 1.8 m/sec (6 ft/sec), and inbed average sorbent particle size of 500 μm) in comparison to the cost of uncontrolled conventional boilers. Annual cost is shown as a function of boiler capacity, coal type, SO_2 control level, and sorbent reactivity. The range of Ca/S ratios listed are from Table 22, developed as shown in Section 7.0.

Table 34 lists the annual cost of AFBC and conventional boilers in terms of $\$/10^6$ Btu output, accounting for the effect of boiler efficiency on system cost. The AFBC costs are summarized in Figures 30 through 32.*

The figures show fixed annual costs, annual operating costs, and total annual costs (the sum of the initial two costs) and represent use of a sorbent with average reactivity. Error bands are included for fixed and total annual costs to illustrate the effect of the estimated accuracy in capital cost estimates of ±30 percent, which is generally the limit for budget equipment estimates. This range is conservative considering that some of the FBC equipment and installation components were estimated based on PEDCo cost data. Therefore, an inaccuracy would be duplicated for certain pieces of equipment in conventional and FBC systems, and the relative comparison of the two technologies would not be affected by these inaccuracies.

Since many of the direct operating costs have been estimated equal for the two technologies (e.g., coal purchase, the unit cost of solid waste disposal, labor, maintenance overhead, chemicals, and process water), no error bands have been assigned to annual operating cost. The conservative estimate of accuracy

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*Although continuous curves are shown, interpolation to other capacities is not recommended. This graphical method was selected to illustrate the economy of scale possible in going to larger boiler capacities.

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TABLE 33. ANNUAL COST OF INDUSTRIAL FBC BOILERS WITH SO_2 CONTROL, DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY=MW 8.8 CONVENTIONAL	8.8 AFBC	22 CONVENTIONAL	22 AFBC	44 CONVENTIONAL	44 AFBC	58.6 CONVENTIONAL	58.6 AFBC
EASTERN HIGH SULFUR (3.54 S)	S 90%	AVERAGE	3.5	927071.	995651.	1826049.	2259707.	3044177.	3856745.	4035712.	[illegible]
		LOW	4.2	927071.	1023547.	1826049.	2340775.	3044177.	4004221.	4035712.	5157161.
		HIGH	2.5	927071.	964394.	1826049.	2185063.	3044177.	3692884.	4035712.	[illegible]
	I 85%	AVERAGE	2.9	927071.	981484.	1826049.	2227139.	3044177.	3785052.	4035712.	[illegible]
		LOW	3.8	927071.	1009982.	1826049.	2306405.	3044177.	3932527.	4035712.	[illegible]
		HIGH	2.1	927071.	956410.	1826049.	2166993.	3044177.	3653962.	4035712.	[illegible]
	M 78.7%	AVERAGE	2.5	927071.	967958.	1826049.	2193515.	3044177.	3711758.	4035712.	4767212.
		LOW	3.4	927071.	996059.	1826049.	2260512.	3044177.	3859254.	4035712.	[illegible]
		HIGH	1.8	927071.	945947.	1826049.	2140444.	3044177.	3597055.	4035712.	[illegible]
	SIP 50%	AVERAGE	1.0	927071.	915153.	1826049.	2065287.	3044177.	3422715.	4035712.	[illegible]
		LOW	1.2	927071.	921503.	1826049.	2080833.	3044177.	3452708.	4035712.	4438527.
		HIGH	0.8	927071.	908792.	1826049.	2049673.	3044177.	3392448.	4035712.	[illegible]
EASTERN LOW SULFUR (0.94 S)	S/I 85.9%	AVERAGE	2.8	921109.	905827.	1836259.	2067308.	3090555.	3437874.	4146445.	4413526.
		LOW	3.7	921109.	912074.	1836259.	2082734.	3090555.	3467890.	4146445.	[illegible]
		HIGH	2.0	921109.	900266.	1836259.	2055540.	3090555.	3410594.	4146445.	4377625.
	M 75%	AVERAGE	2.2	921109.	901188.	1836259.	2055822.	3090555.	3415082.	4146445.	[illegible]
		LOW	3.2	921109.	908136.	1836259.	2073009.	3090555.	3448842.	4146445.	[illegible]
		HIGH	1.6	921109.	897012.	1836259.	2045471.	3090555.	3394870.	4146445.	[illegible]
SUBBITUMINOUS LOW SULFUR (0.64 S)	S/I 83.2%	AVERAGE	2.7	955882.	864712.	1763479.	1905824.	3028267.	3095923.	3999383.	[illegible]
		LOW	3.6	955882.	870700.	1763479.	1920617.	3028267.	3124918.	3999383.	[illegible]
		HIGH	2.0	955882.	860048.	1763479.	1894277.	3028267.	3073204.	3999383.	[illegible]
	M 75%	AVERAGE	2.2	955882.	860967.	1763479.	1896552.	3028267.	3077680.	3999383.	[illegible]
		LOW	3.2	955882.	867620.	1763479.	1913024.	3028267.	3110038.	3999383.	[illegible]
		HIGH	1.6	955882.	856966.	1763479.	1888633.	3028267.	3058122.	3999383.	[illegible]

Note: Conventional Boilers shown here contain no provisions for SO_2 control. This comparison was made according to the groundrules of the industrial boiler system study. If costs of SO_2 control are included AFBC becomes more competitive.

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TABLE 34. ANNUAL COST ($/$10^6$ Btu OUTPUT) OF INDUSTRIAL FBC BOILERS WITH SO_2 CONTROL

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY—MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	7.59	7.75	5.76	6.96	4.77	5.91	4.58	5.69
		LOW	4.2	7.59	8.04	5.76	7.28	4.77	6.17	4.58	5.97
		HIGH	2.3	7.59	7.42	5.76	6.66	4.77	5.66	4.58	5.38
	I 85%	AVERAGE	2.9	7.59	7.62	5.76	6.84	4.77	5.78	4.58	5.56
		LOW	3.8	7.59	7.91	5.76	7.15	4.77	6.06	4.58	5.84
		HIGH	2.1	7.59	7.38	5.76	6.60	4.77	5.53	4.58	5.31
	M 78.7%	AVERAGE	2.5	7.59	7.48	5.76	6.72	4.77	5.65	4.58	5.43
		LOW	3.4	7.59	7.78	5.76	6.99	4.77	5.93	4.58	5.71
		HIGH	1.8	7.59	7.20	5.76	6.51	4.77	5.40	4.58	5.22
	SIP 56%	AVERAGE	1.0	7.59	7.00	5.76	6.25	4.77	5.15	4.58	4.95
		LOW	1.2	7.59	7.06	5.76	6.31	4.77	5.21	4.58	5.01
		HIGH	0.8	7.59	6.95	5.76	6.19	4.77	5.10	4.58	4.90
EASTERN LOW SULFUR (0.9% S)	S/I 85.9%	AVERAGE	2.8	7.12	6.87	5.62	6.21	4.70	5.13	4.55	4.93
		LOW	3.7	7.12	6.93	5.62	6.27	4.70	5.19	4.55	4.98
		HIGH	2.0	7.12	6.82	5.62	6.16	4.70	5.08	4.55	4.89
	M 75%	AVERAGE	2.2	7.12	6.85	5.62	6.17	4.70	5.10	4.55	4.91
		LOW	3.2	7.12	6.90	5.62	6.24	4.70	5.16	4.55	4.96
		HIGH	1.6	7.12	6.79	5.62	6.13	4.70	5.06	4.55	4.86
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 85.2%	AVERAGE	2.7	7.41	6.75	5.54	5.88	4.73	4.75	4.57	4.51
		LOW	3.6	7.41	6.79	5.54	5.93	4.73	4.81	4.57	4.56
		HIGH	2.0	7.41	6.69	5.54	5.83	4.73	4.71	4.57	4.47
	M 75%	AVERAGE	2.2	7.41	6.70	5.54	5.84	4.73	4.71	4.57	4.48
		LOW	3.2	7.41	6.77	5.54	5.91	4.73	4.77	4.57	4.55
		HIGH	1.6	7.41	6.66	5.54	5.80	4.73	4.66	4.57	4.44

Note: Conventional Boilers shown here contain no provisions for SO_2 control. This comparison was made according to the groundrules of the industrial boiler system study. If costs of SO_2 control are included AFBC becomes more competitive.

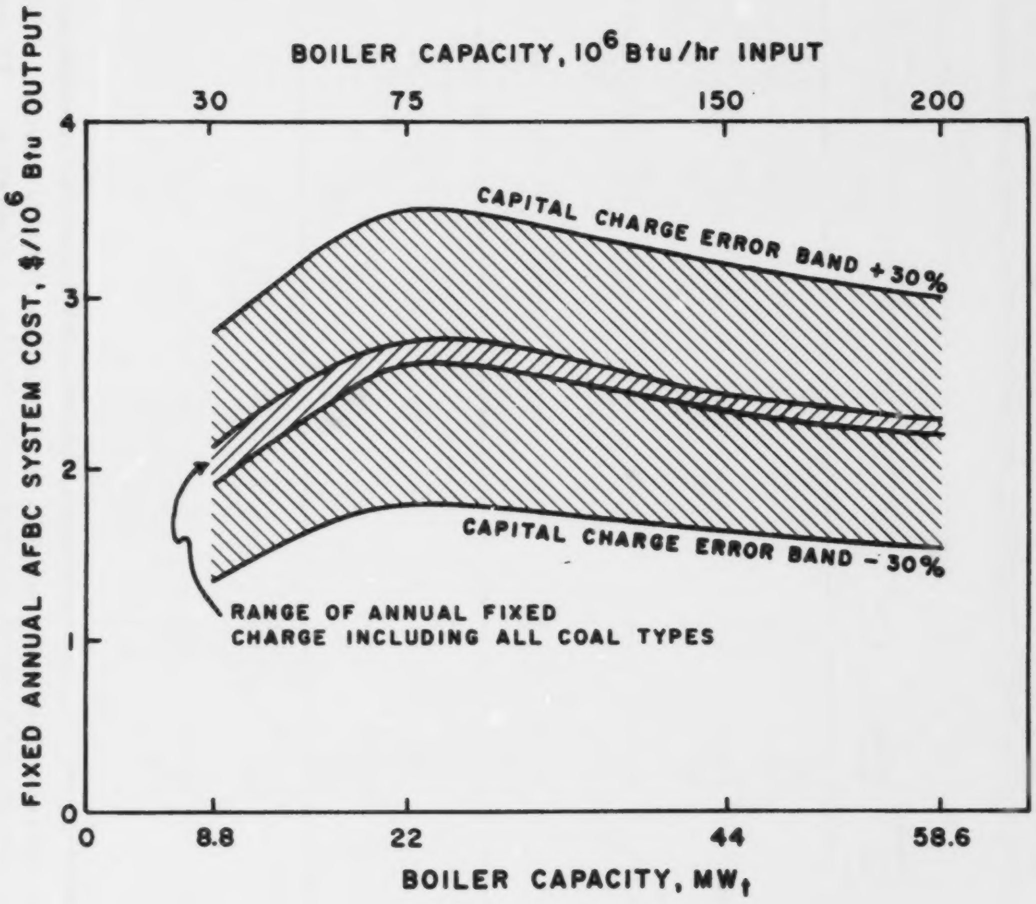

Figure 30. Annual fixed charge of FBC with SO_2 control.

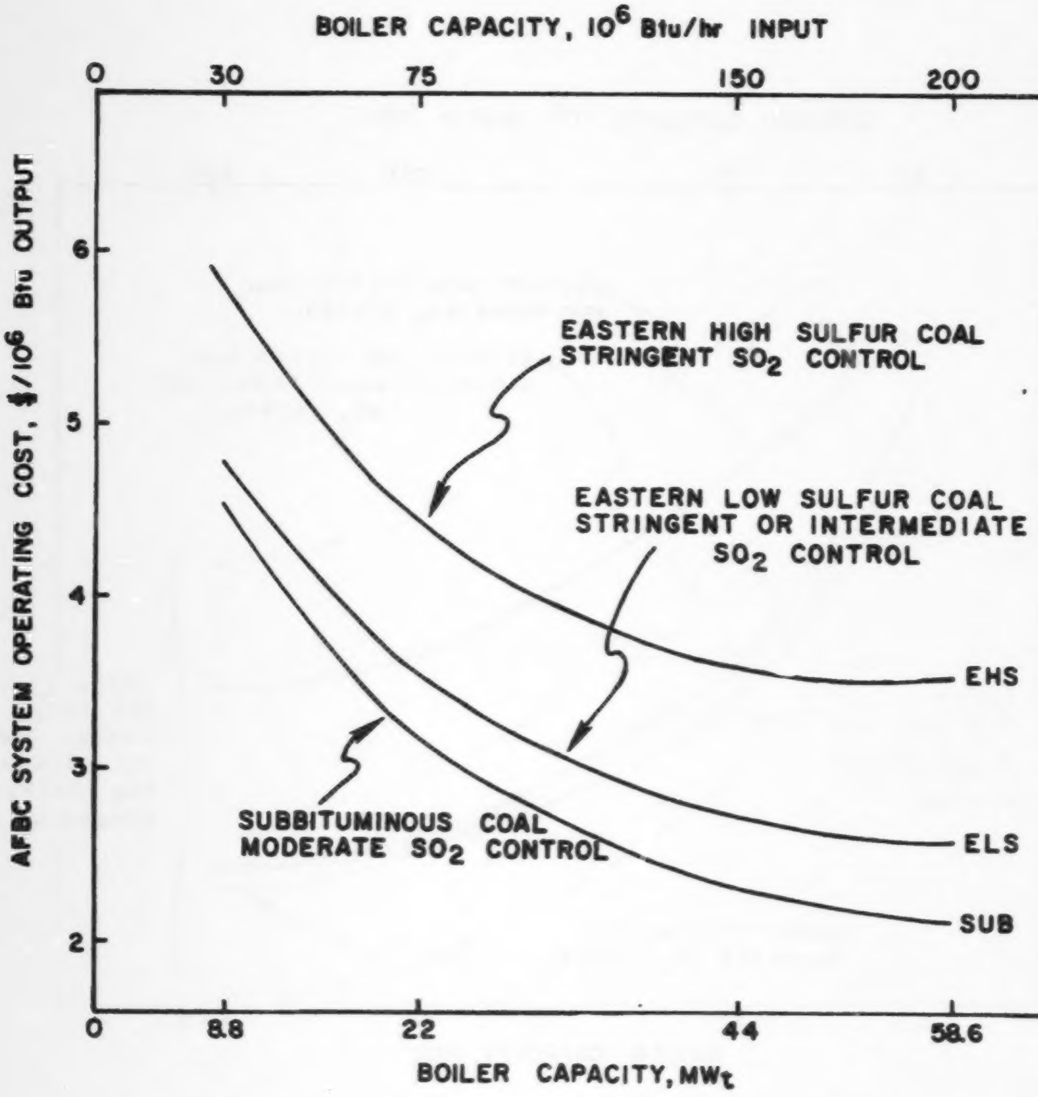

Figure 31. Total operating cost of AFBC with SO_2 control.

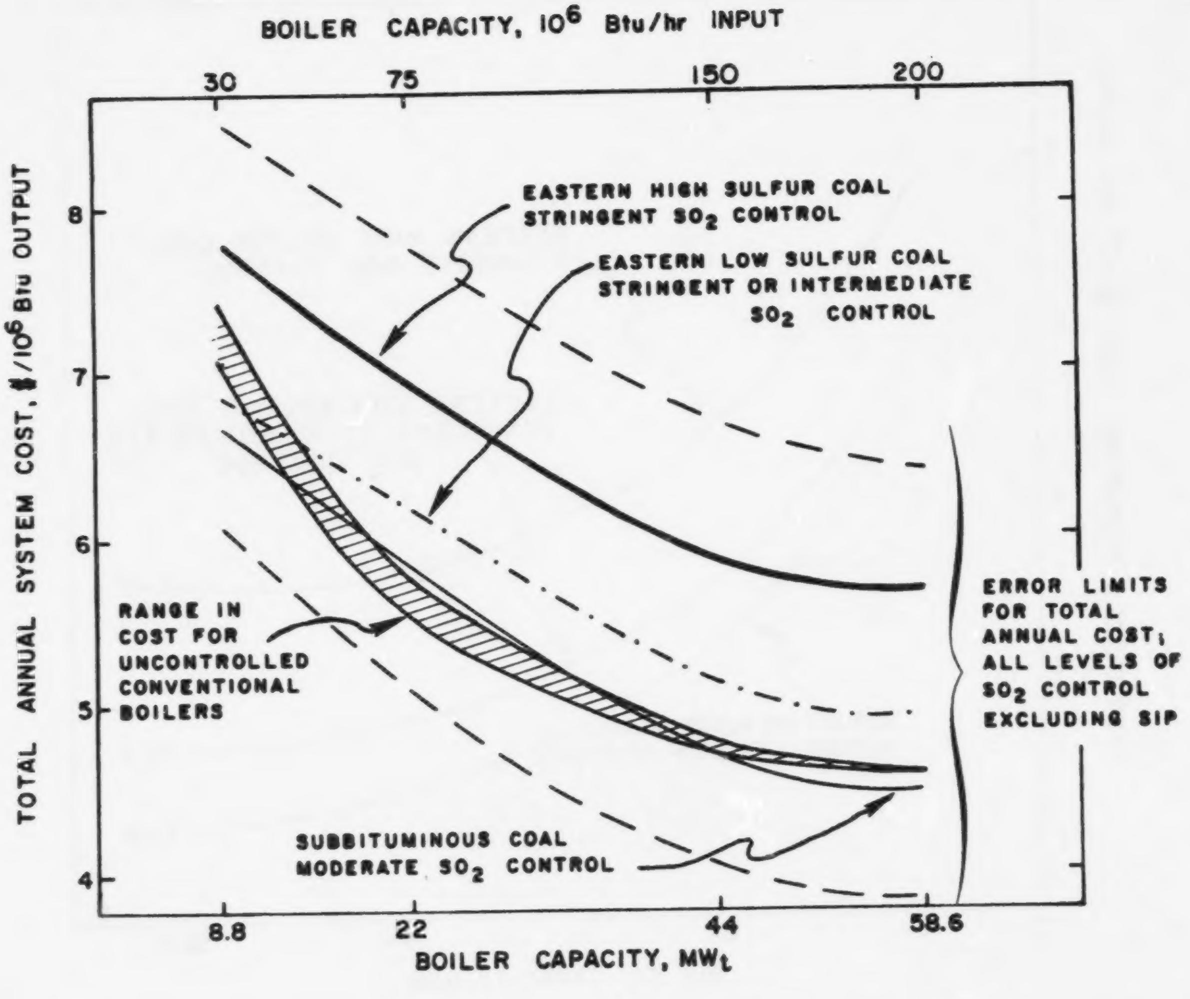

Figure 32. Total annual cost of FBC with SO_2 control.

assumed for the capital costs should account for the possible errors in limestone purchase or electricity requirements. Therefore, the error bands shown on the graph of fixed annual cost (capital charges) have been translated directly to the graph of total annual cost. The error band heights are 0.70 to 0.80 $\$/10^6$ Btu, both positive and negative for total annual FBC cost. This is about 9 percent of total cost for the 8.8 MW_t AFBC boiler burning Eastern high sulfur coal at a stringent SO_2 control level, and increases to a maximum of 15 percent for the 58.6 MW_t AFBC boiler burning subbituminous coal.

System operating costs drop off significantly as boiler capacity increases due partially to increasing boiler efficiency. A second reason is the underlying cost of labor and overhead. These costs do not increase directly with boiler capacity since there is some minimum staffing and overhead requirement for the small boiler capacity which increases slowly in proportion to boiler capacity.

Two conclusions are drawn from the graph of total annual cost. First, FBC with stringent SO_2 control firing high sulfur coal is about 20 percent more expensive that use of uncontrolled conventional boilers. The only exception is in the cost of the 8.8 MW_t boiler where system costs are very similar, the AFBC is only about 7 percent more costly than the uncontrolled conventional boiler. The small AFBC boiler has a relatively low capital cost because of its simple, space saving, package design. It is based on a cost quote from one vendor that is starting to penetrate the commercial market. However, it can be argued that the cost is slightly underestimated for the purpose of marketing. If the 30 percent capital error band is factored in and the conventional costs are assumed to be accurate, the maximum added total cost of the controlled 8.8 MW_t FBC over the uncontrolled conventional system is 15 percent. The actual cost differential probably falls within the range of 5 to 15 percent.

The second conclusion is that the difference in cost between stringent SO_2 control with high sulfur coal and moderate SO_2 control with subbituminous coal is roughly 1.10 to 1.40 $\$/10^6$ Btu output. If the error bands are considered, this margin widens to a maximum of 2.70 $\$/10^6$ Btu.

4.3.4 "Commercially-Offered" AFBC Industrial Boilers Versus "Best Systems" of SO_2 Control

Atmospheric fluidized-bed combustion is am emerging technology, and design/operating parameters currently specified by FBC vendors are generally different than those specified in this report for "best system" design. The principal differences relating to SO_2 control performance include gas phase residence time and sorbent particle size in the bed.

It is important to note that the phrase "best system" as used in this report, refers to design/operating conditions selected by GCA to minimize sorbent use, spent solids generation, and provide the low cost approach to controlling SO_2 in FBC. It is not intended to denote that one vendor's design is superior to another or that current technology is far removed from the recommended conditions.

The increase to "best system" gas phase residence time of 0.67 sec (commercial systems operate at roughly 0.5 sec and below) can be achieved by using either deeper beds or lower gas velocities. Deeper beds require increased furnace height while lower gas velocities require larger furnace cross section. Either modification is achieved at the expense of increased capital investment.

Although these capital cost increases would result, the additional expenditure may be recovered by reduced capital requirements elsewhere. For instance, increasing the bed depth sufficiently will allow design of natural circulation boilers instead of forced convection boilers, thus eliminating recirculation pumps. Shallow beds do not allow enough slope in the steam tubes for natural convection. In addition, lower superficial velocities may result in a capital cost savings in primary particulate control through reduced bed elutriation.

It is expected that lower operating costs should recover any added capital expenditure associated with increased gas residence time. Reduced sorbent and coal requirements, lower maintenance through elimination of convection pumps, and less abrasion of internals such as steam tubes and cyclones are all areas where savings could occur. Except for reduced sorbent cost, these savings cannot be quantified until AFBC is demonstrated in commercial operation.

As mentioned, Westinghouse Research and Development is conducting an engineering evaluation of industrial fluidized-bed technology. The study will assess the net cost and performance impact of implementing "best system" design/operating conditions.

The Babcock & Wilcox Company has estimated the effect of three different superficial velocities on cost in terms of $/kW.[33] As illustrated by Figure 33, the lowest velocity results in the lowest capital cost, coal cost, and limestone cost. Although these estimates were developed for utility applications, the relative proportions should hold true for industrial applications.

Reduction of limestone particle size to the recommended in-bed average of 500 μm can provide similar, and possibly greater, operating cost savings than noted for the lengthening of gas residence time. The data presented in Sections 2.0 and 3.0 illustrate that limestone consumption can be reduced by at least 20 percent if average in-bed sorbent particle size is reduced from 1,000 μm to 500 μm. The difference is even greater for sorbents of low reactivity. In several cases, vendors are specifying limestone feed particle sizes of greater than 1,000 μm, possibly as high as 1,500 μm. One uncertainty is that the relationship between feed sorbent size, and the actual size that exists

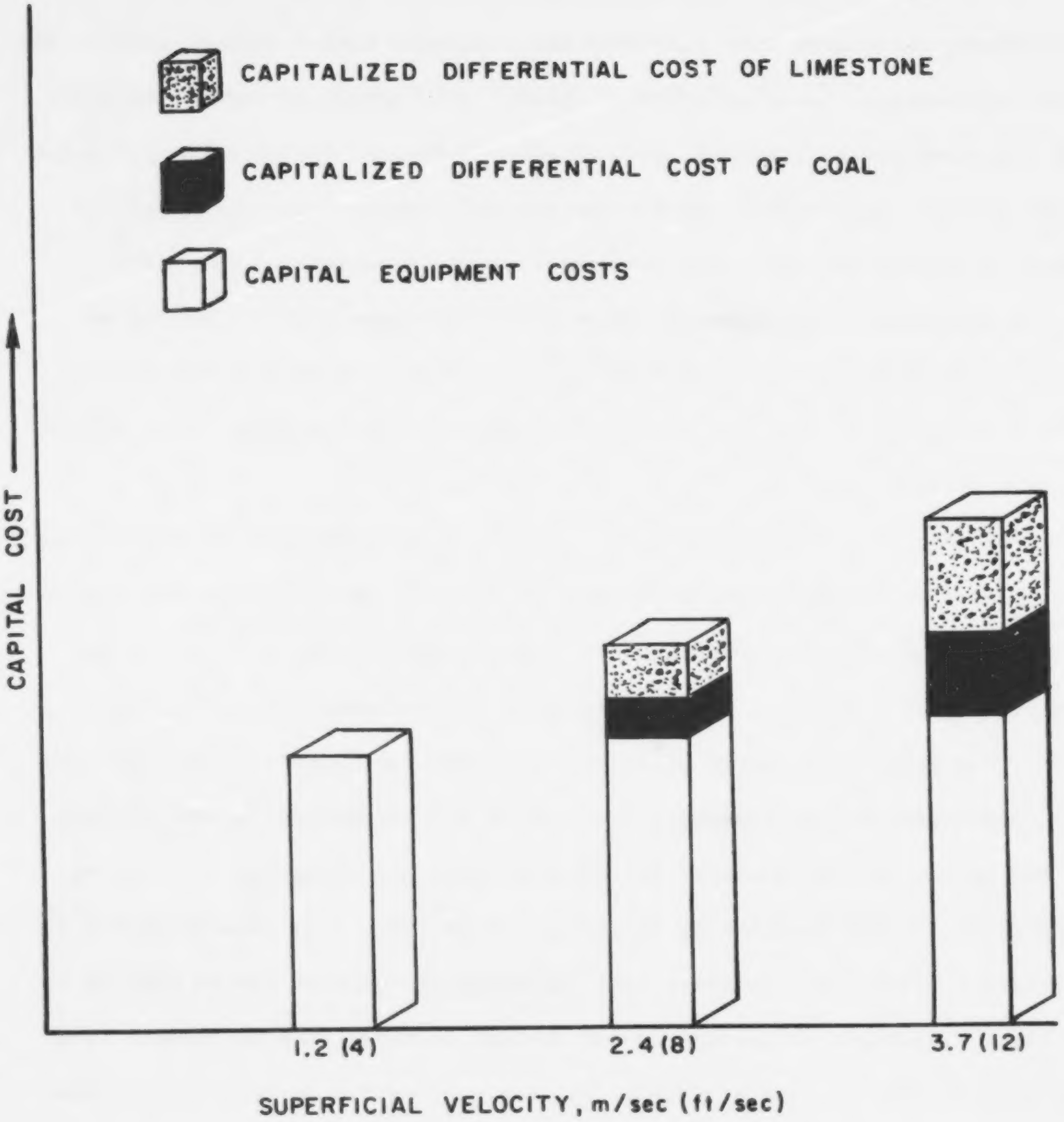

Figure 33. FBC cost variation as a function of superficial velocity, after Babcock and Wilcox.

in the bed, is not rigorously predictable. Therefore, although vendors may be quoting mass mean particle sizes of 1,000 to 1,500 μm for the sorbent feed, the actual size which might exist in the bed could be much closer to the 500 μm surface mean which is considered for the "best" system.

Incremental costs which could result from using smaller sorbent particles include the added unit cost of sorbent (in the form of increased purchase cost or onsite screening facilities), any added cost of primary particle control, and any added maintenance requirements.

If particle size is reduced, there may be savings in the cost of primary and final particulate control equipment since the amount of sorbent is decreasing at the same time that the proportion of elutriated bed material is rising. Westinghouse has formulated projections of elutriated solids loadings from atmospheric FBC as a function of Ca/S ratio based on Greer limestone.[34] In the atmospheric case, lowering the Ca/S ratio from 5 to 2 resulted in a 45 percent reduction in solids elutriated from the bed. This implies that fine particle elutriation can increase (as a result of particle size reduction to reduce sorbent needs) some measurable amount before the cost of primary particulate control increases significantly. Further experimentation is required to determine where this breakpoint exists.

Any significantly increased maintenance costs resulting from sorbent size reduction would be in the form of replacement part costs for abraded internal equipment. The magnitude of this added cost, however, is anticipated to be small in comparison to overall plant cost, but must be confirmed in commercial operation.

Again, it is emphasized that, in order to maintain a surface mean sorbent size of 500 μm in the bed, it may not always be necessary to reduce feed sorbent particle size significantly from the 1,000 to 1,500 μm mass mean feed size specified by many vendors.

Based on the foregoing discussion, the most readily quantifiable difference in cost between "commercially offered" and "best system" is sorbent purchase. The cost of "commercially-offered" systems is estimated here using the results of the cost sensitivity analysis and estimates of sorbent requirements based on the Westinghouse SO_2 kinetic model. Changes in Ca/S ratio, as projected using the Westinghouse model, are reported in Section 3.0 considering the "commercially-offered" systems and several different sorbents. As shown in Table 21, significant reductions in required Ca/S ratio may be possible by increasing gas residence time and reducing sorbent particle size, according to the model projections.

Depending on limestone type and reactivity and SO_2 control level, Ca/S ratios are noted to rise to above 10 in Table 21, a value which would not be used in practice. A Ca/S ratio over 6 or 7 may be economically uncompetitive due to added operating cost (see Figure 34) and losses in boiler efficiency.*

The effect of increasing Ca/S ratio on annual operating cost is shown in Figure 34. A detailed discussion of the method of calculation is given in the sensitivity analysis (Subsection 4.3.8). Briefly, cost estimates were prepared for "best systems" using the cost basis described previously. Then, baseline design/operating conditions were selected (see Table C-3) and single parameters, such as Ca/S ratio, were varied individually to assess their impact on FBC system cost. The sensitivity cost curves for Ca/S ratio are linear as shown in Figure 34, and show an added cost between 35 to 37 ¢/10^6 Btu output for

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*The energy sensitivity analysis presented in Section 5.5 of this report indicates that as Ca/S ratio exceeds a value in the range of 5.5 to 6.0, the efficiency of an AFBC boiler drops below that of an uncontrolled stoker (see Figure 52).

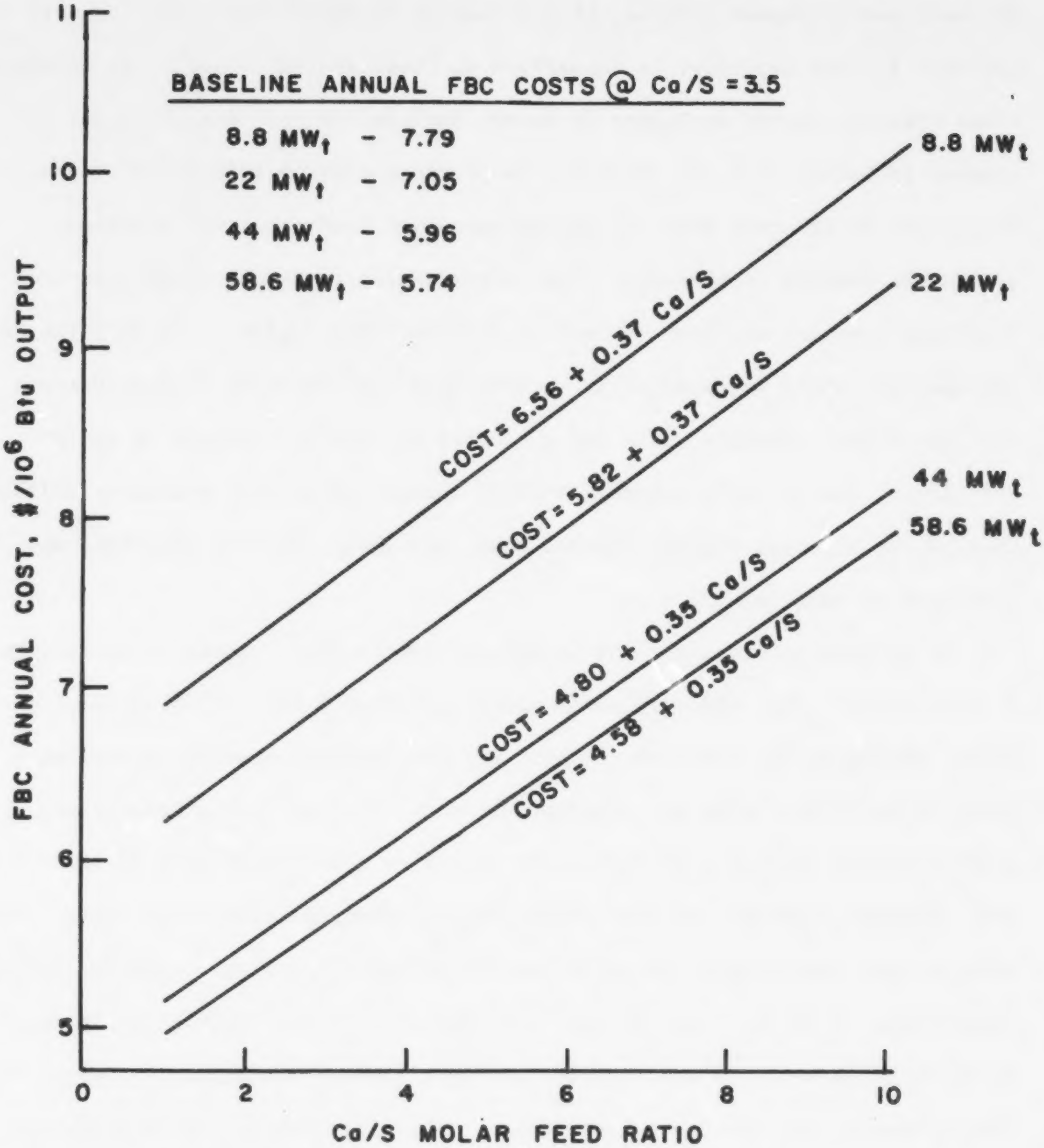

Figure 34. AFBC cost as a function of Ca/S molar feed ratio. (All other design/operating parameters constant).

incremental increases of one in Ca/S ratio. If Ca/S ratios as high as 6 or 7 are used for stringent control, then a loss of $1.00/$10^6$ Btu output or more is incurred for any capacity, in comparison to "best system" design. It is emphasized that the curves in Figure 34 do not include any cost penalties for increases in boiler size, or increases in particle control requirements, etc., that might be incurred when FBC design operating conditions are adjusted to obtain the reduced Ca/S levels, since other design/operating conditions have been held constant at "best system" conditions. The figure shows only the cost savings that could be expected if sorbent feed rate could be reduced without such penalties. Changes which are accounted for include changes in boiler efficiency, yearly sorbent purchase cost, annualized sorbent and spent solids storage costs, spent solids disposal cost, and power cost for limestone and spent solids handling.

As an example, consider the Georgetown AFBC boiler. Table 21 in Section 3.0 illustrates that the Ca/S ratio might fall from 5.29 to 2.85 (Greer limestone, stringent SO_2 control) if "best system" design/operating conditions are used, as projected using the Westinghouse model. Based on the sensitivity analysis shown in Figure 34 this could result in a cost reduction of $0.90/$10^6$ Btu. However, considering the uncertainty of other operating and capital cost changes that could result in adjusting the Georgetown design to achieve "best" conditions, it is fair only to conclude that these other costs could increase by $0.90/$10^6$ Btu before modification to "best system" conditions would not be cost effective for the Georgetown design. A similar conclusion could be drawn for other specific designs and limestone types.

As an added point, consider the 8.8 MW_t FBC boiler burning high sulfur coal. A 20 percent reduction in limestone use would reduce annual purchase cost by roughly \$4,000 for any control level other than the average SIP level of 56 percent. This operating savings can be translated to an increased capital cost allowance. The present worth of this annual cost over 30 years (to be consistent with boiler life expectancy) at an interest rate of 10 percent is \$38,000. This value is 15 percent of the cost of boiler equipment for the 8.8 MW_t boiler and indicates the approximate capital equipment cost increase which can be accommodated with no concomitant increase in the annual cost of steam production because of operating cost savings.

4.3.5 Cost Comparison: AFBC "Best System" Designs Versus Conventional Boilers Without SO_2 Emission Control

The goal of this cost study is to compare the total cost of controlled FBC with uncontrolled conventional boilers so that the incremental cost of using FBC as a boiler system controlling SO_2 emissions can be isolated. Similar documents are being prepared by other contractors to estimate the cost of other SO_2 control technologies with the same conventional boiler costs as a basis. These technologies include flue gas desulfurization, coal cleaning, oil cleaning, and synthetic fuels. A future study by EPA will compare the cost of SO_2 removal using FBC and the other technologies based on these documents.

The preceding subsection introduced the comparison of uncontrolled conventional and controlled FBC industrial boiler cost. The intent of this subsection is to present a more detailed analysis indicating cost differences which exist as a function of sorbent reactivity, SO_2 control level, and coal type. These data are itemized in Tables 35 through 38 and are depicted graphically in Figures 35 through 37. Tabulated costs are shown in terms of $\$/10^6$ Btu output, \$/J/sec thermal input, and $\$/10^6$ Btu/hr thermal input. The latter two cost parameters are shown for consistency with guidelines established for all of the technology assessment reports. The costs reported graphically are in terms of $\$/10^6$ Btu

TABLE 35. COSTS OF "BEST" SO_2 CONTROL TECHNIQUES FOR COAL-FIRED AFBC BOILERS OF 8.8 MW_t (30 × 10^6 Btu/hr) CAPACITY

AFBC with SO_2 control				Total annualized costs			Impacts	
Standard boiler capacity MW_t (10^6 Btu/hr) and coal characteristics	SO_2 control level and percentage reduction	Sorbent reactivity	Ca/S ratio	\$/10^6 Btu output	\$/J/sec thermal input	\$/10^6 Btu/hr thermal input	Percent increase in costs over uncontrolled conventional boilers*	Percent increase in costs over SIP controlled AFBC boilers
8.8 (30) Eastern high sulfur coal (3.5% S)	S 90%	Average	3.3	7.75	0.113	33,200	7.4	8.8
		Low	4.2	8.04	0.116	34,100	10.4	11.1
		High	2.3	7.42	0.110	32,100	4.0	6.1
	I 85%	Average	2.9	7.62	0.112	32,700	5.9	7.3
		Low	3.8	7.91	0.115	33,700	8.9	9.6
		High	2.1	7.36	0.109	31,900	3.2	5.3
	M 78.7%	Average	2.5	7.48	0.110	32,300	4.4	5.8
		Low	3.4	7.78	0.113	33,200	7.4	8.1
		High	1.8	7.26	0.107	31,500	2.0	4.1
	SIP 56%	Average	1.0	7.00	0.104	30,500	-1.3	-
		Low	1.2	7.06	0.105	30,700	-0.6	-
		High	0.8	6.93	0.103	30,300	-2.0	-
8.8 (30) Eastern low sulfur coal (0.9% S)	S or I 83.9%	Average	2.8	6.87	0.103	30,200	-1.7	-
		Low	3.7	6.93	0.104	30,400	-1.0	-
		High	2.0	6.82	0.102	30,000	-2.3	-
	M 75%	Average	2.2	6.83	0.102	30,000	-2.2	-
		Low	3.2	6.90	0.103	30,300	-1.4	-
		High	1.6	6.79	0.102	29,900	-2.6	-
8.8 (30) Subbituminous coal (0.6% S)	S or I 83.2%	Average	2.7	6.73	0.098	28,800	-7.4	-
		Low	3.6	6.79	0.099	29,000	-6.8	-
		High	2.0	6.69	0.098	28,700	-7.9	-
	M 75%	Average	2.2	6.70	0.098	28,700	-7.8	-
		Low	3.2	6.77	0.099	28,900	-7.1	-
		High	1.6	6.66	0.097	28,600	-8.2	-

*Based on costs in terms of \$/J/sec (\$/10^6 Btu/hr) thermal input.

TABLE 36. COSTS OF "BEST" SO_2 CONTROL TECHNIQUES FOR COAL-FIRED AFBC BOILERS OF 22 MW_t (75 × 10^6 Btu/hr) CAPACITY

AFBC with SO_2 control				Total annualized costs			Impacts	
Standard boiler capacity MW_t (10^6 Btu/hr) and coal characteristics	SO_2 control level and percentage reduction	Sorbent reactivity	Ca/S ratio	$/$10^6$ Btu output	$/J/sec thermal input	$/$10^6$ Btu/hr thermal input	Percent increase in costs over uncontrolled conventional boilers*	Percent increase in costs over SIP controlled AFBC boilers
22 (75) Eastern high sulfur coal (3.5% S)	S 90%	Average	3.3	6.96	0.103	30,100	23.7	9.4
		Low	4.2	7.28	0.106	31,200	28.2	12.5
		High	2.3	6.66	0.099	29,100	19.7	6.6
	I 85%	Average	2.9	6.84	0.101	29,700	22.0	7.8
		Low	3.8	7.15	0.105	30,800	26.3	10.8
		High	2.1	6.60	0.099	28,900	18.7	5.7
	M 78.7%	Average	2.5	6.72	0.100	29,200	20.1	6.2
		Low	3.4	6.99	0.103	30,100	23.8	8.6
		High	1.8	6.51	0.097	28,500	17.2	4.4
	SIP 56%	Average	1.0	6.25	0.094	27,500	13.1	-
		Low	1.2	6.31	0.095	27,700	14.0	-
		High	0.8	6.19	0.093	27,300	12.2	-
22 (75) Eastern low sulfur coal (0.9% S)	S or I 83.9%	Average	2.8	6.21	0.094	27,600	12.6	-
		Low	3.7	6.27	0.095	27,800	13.4	-
		High	2.0	6.16	0.093	27,400	11.8	-
	M 75%	Average	2.2	6.17	0.093	27,400	12.0	-
		Low	3.2	6.24	0.094	27,600	12.9	-
		High	1.6	6.13	0.093	27,300	11.4	-
22 (75) Subbituminous coal (0.6% S)	S or I 83.2%	Average	2.7	5.88	0.087	25,400	8.1	-
		Low	3.6	5.93	0.087	25,600	8.9	-
		High	2.0	5.83	0.086	25,300	7.4	-
	M 75%	Average	2.2	5.84	0.086	25,300	7.6	-
		Low	3.2	5.91	0.087	25,500	8.5	-
		High	1.6	5.80	0.086	25,200	7.0	-

*Based on costs in terms of $/J/sec ($/10^6 Btu/hr) thermal input.

TABLE 37. COSTS OF "BEST" SO_2 CONTROL TECHNIQUES FOR COAL-FIRED AFBC BOILERS OF 44 MW_t (150 × 10^6 Btu/hr) CAPACITY

AFBC with SO_2 control				Total annualized costs			Impacts	
Standard boiler capacity MWt (10^6 Btu/hr) and coal characteristics	SO_2 control level and percentage reduction	Sorbent reactivity	Ca/S ratio	\$/10^6 Btu output	\$/J/sec thermal input	\$/10^6 Btu/hr thermal input	Percent increase in costs over uncontrolled conventional boilers*	Percent increase in costs over SIP controlled AFBC boilers
44 (150) Eastern high sulfur coal (3.5% S)	S 90%	Average	3.3	5.91	0.088	25,700	26.7	12.7
		Low	4.2	6.19	0.091	26,700	31.5	16.0
		High	2.3	5.60	0.084	24,600	21.3	8.9
	I 85%	Average	2.9	5.78	0.086	25,200	24.3	10.6
		Low	3.8	6.06	0.089	26,200	29.2	13.9
		High	2.1	5.53	0.083	24,400	20.0	7.7
	M 78.7%	Average	2.5	5.65	0.084	24,700	21.9	8.4
		Low	3.4	5.93	0.088	25,700	26.8	11.8
		High	1.8	5.44	0.082	24,000	18.2	6.0
	SIP 56%	Average	1.0	5.15	0.078	22,800	12.4	-
		Low	1.2	5.21	0.079	23,000	13.4	-
		High	0.8	5.10	0.077	22,600	11.4	-
44 (150) Eastern low sulfur coal (0.9% S)	S or I 83.9%	Average	2.8	5.13	0.078	22,900	11.2	-
		Low	3.7	5.19	0.079	23,100	12.2	-
		High	2.0	5.08	0.078	22,700	10.4	-
	M 75%	Average	2.2	5.10	0.078	22,800	10.5	-
		Low	3.2	5.16	0.078	23,000	11.6	-
		High	1.6	5.06	0.077	22,600	9.8	-
44 (150) Subbituminous coal (0.6% S)	S or I 83.2%	Average	2.7	4.75	0.070	20,600	2.2	-
		Low	3.6	4.80	0.071	20,800	3.2	-
		High	2.0	4.70	0.070	20,500	1.5	-
	M 75%	Average	2.2	4.71	0.070	20,500	1.6	-
		Low	3.2	4.77	0.071	20,700	2.7	-
		High	1.6	4.68	0.070	20,400	1.0	-

*Based on costs in terms of \$J/sec (\$/10^6 Btu/hr) thermal input.

TABLE 38. COSTS OF "BEST" SO_2 CONTROL TECHNIQUES FOR COAL-FIRED AFBC BOILERS OF 58.6 MW_t (200 × 10^6 Btu/hr) CAPACITY

AFBC with SO_2 control				Total annualized costs			Impacts	
Standard boiler capacity MW_t (10^6 Btu/hr) and coal characteristics	SO_2 control level and percentage reduction	Sorbent reactivity	Ca/S ratio	$/$10^6$ Btu output	$/J/sec thermal input	$/$10^6$ Btu/hr thermal input	Percent increase in costs over uncontrolled conventional boilers*	Percent increase in costs over SIP controlled AFBC boilers
58.6 (200) Eastern high sulfur coal (3.5% S)	S 90%	Average	3.3	5.69	0.085	24,800	22.9	12.9
		Low	4.2	5.97	0.088	25,800	27.8	16.2
		High	2.3	5.38	0.081	23,700	17.5	8.9
	I 85%	Average	2.9	5.56	0.083	24,300	20.5	10.7
		Low	3.8	5.84	0.086	25,300	25.4	14.1
		High	2.1	5.31	0.080	23,500	16.2	7.7
	M 78.7%	Average	3.4	5.43	0.081	23,800	18.1	8.5
		Low	3.4	5.71	0.085	24,800	23.0	11.9
		High	1.8	5.22	0.079	23,100	14.3	6.0
	SIP 56%	Average	1.0	4.95	0.075	22,000	8.9	-
		Low	1.2	5.01	0.076	22,200	9.9	-
		High	0.8	4.90	0.074	21,800	7.9	-
58.6 (200) Eastern low sulfur coal (0.9% S)	S or I 83.9%	Average	2.8	4.93	0.075	22,100	6.4	-
		Low	3.7	4.99	0.076	22,300	7.4	-
		High	2.0	4.89	0.075	21,900	5.6	-
	M 75%	Average	2.2	4.90	0.075	21,900	5.7	-
		Low	3.2	4.96	0.076	22,100	6.8	-
		High	1.6	4.86	0.074	21,800	5.1	-
58.6 (200) Subbituminous coal (0.6% S)	S or I 83.2%	Average	2.7	4.51	0.067	19,600	-1.8	-
		Low	3.6	4.56	0.068	19,800	-0.8	-
		High	2.0	4.47	0.067	19,500	-2.5	-
	M 75%	Average	2.2	4.48	0.067	19,500	-2.4	-
		Low	3.2	4.54	0.067	19,700	-1.3	-
		High	1.6	4.44	0.066	19,400	-3.0	-

*Based on costs in terms of $/J/sec ($/10^6 Btu/hr) thermal input.

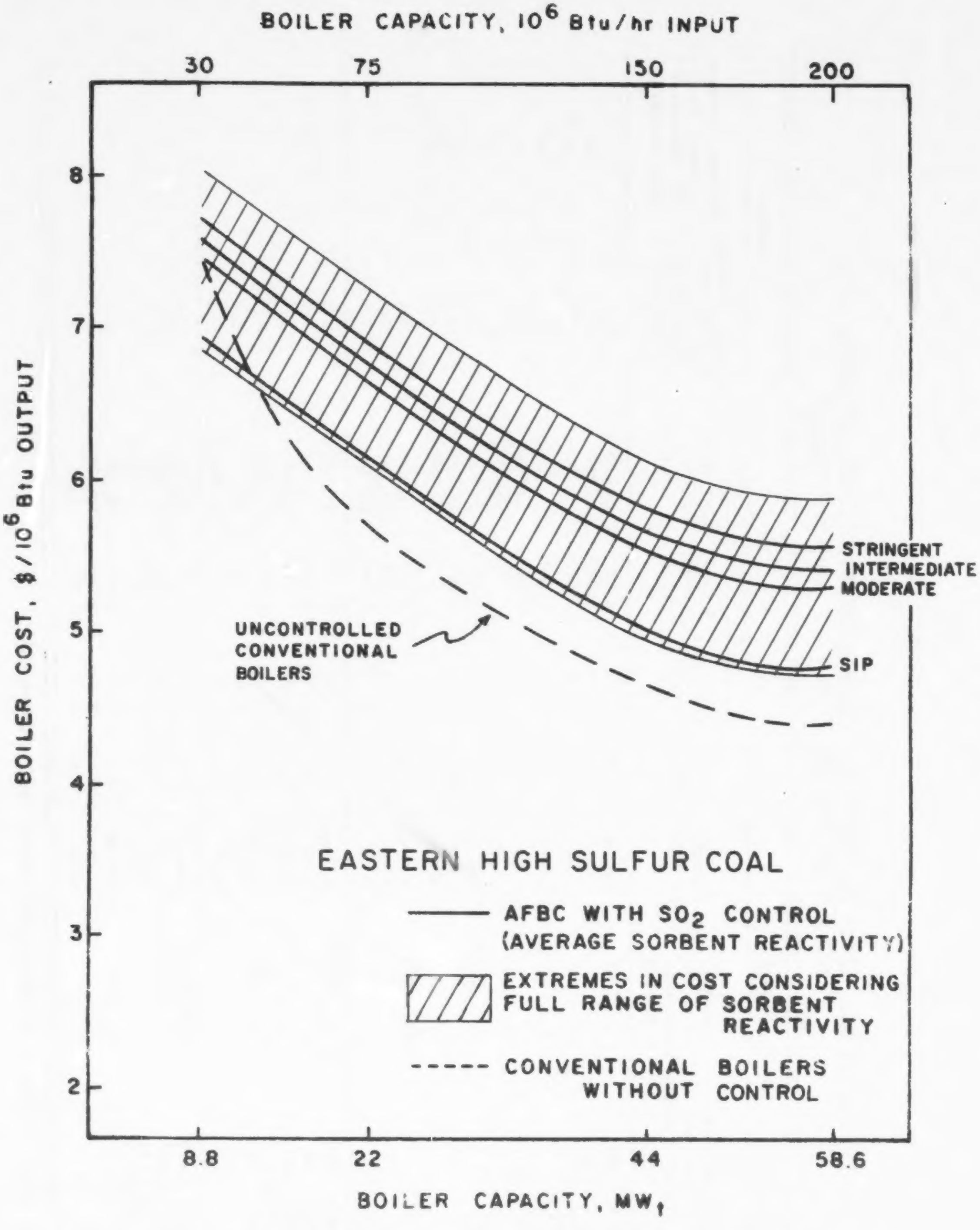

Figure 35. Cost comparison: AFBC boilers with SO_2 control versus uncontrolled conventional boilers; Eastern high sulfur coal.

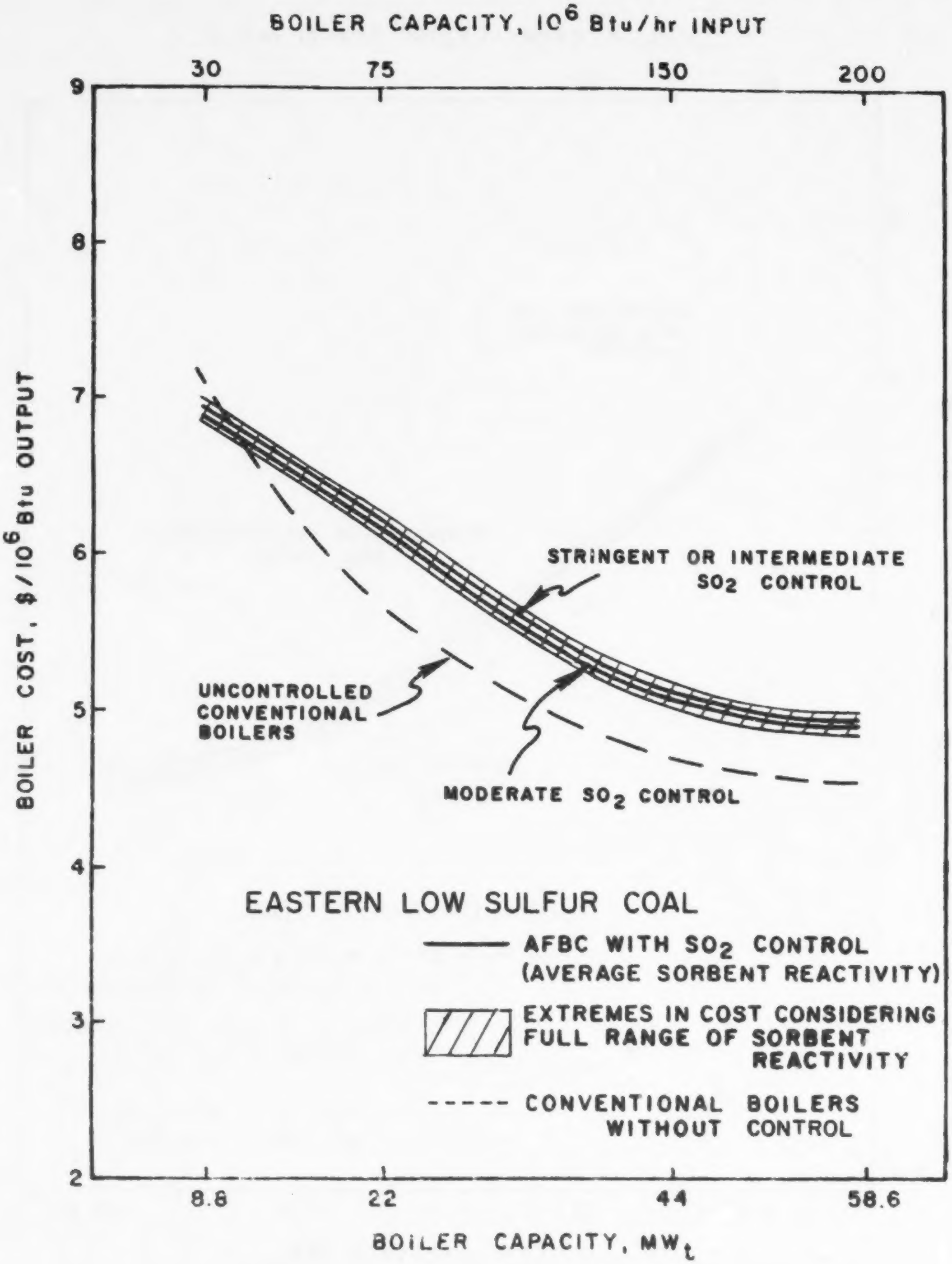

Figure 36. Cost comparison: AFBC boilers with SO_2 control versus uncontrolled conventional boilers; Eastern low sulfur coal.

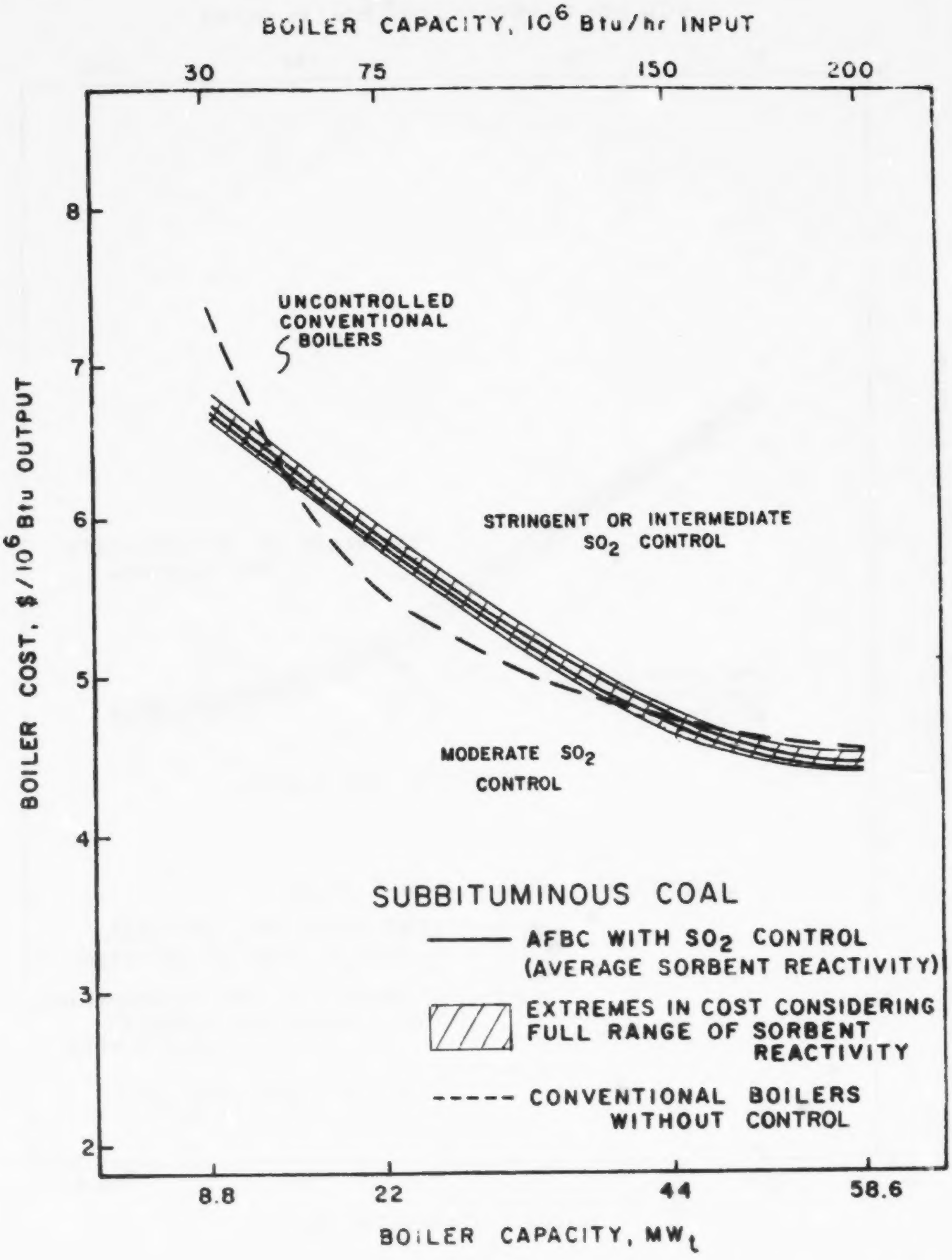

Figure 37. Cost comparison: AFBC boilers with SO_2 control versus uncontrolled conventional boilers; subbituminous coal.

output because it is a more readily interpreted parameter. The cost curves shown for the AFBC boilers represent average sorbent reactivity. Crosshatching is included to indicate extremes in cost based on variation in sorbent reactivity. Depending on the range considered, sorbent reactivity can have more impact on cost than SO_2 control level.

Each figure represents one of the three different coals and illustrates the expected result that FBC industrial boiler technology with SO_2 control is generally more expensive than conventional uncontrolled industrial boiler systems. The greatest difference is noted with high sulfur coal where the most sulfur must be removed in the FBC system. The cost differential becomes smaller with lower sulfur coals. In fact, the controlled 8.8 MW_t FBC boiler has a comparable or even slightly lower total cost than the uncontrolled conventional 8.8 MW_t boiler. This low cost results principally because of the small additional capital and operating cost associated with limestone use and spent solids handling when low sulfur coals are burned in FBC boilers. The difference also may be a function of the use of cost quotes from separate sources. The effect of inaccuracy in boiler cost estimates on total annual system cost is discussed in more detail in the cost sensitivity analysis in Subsection 4.3.8. The influence of estimating error associated with present cost estimates is also seen in Figure 32, shown previously.

Another crossover in cost is demonstrated where the subbituminous coal-fired AFBC boiler is compared with the uncontrolled pulverized coal boiler. The cost similarity is attributed to the technical complexity of the conventional unit and the minimal sulfur removal requirements in the AFBC boiler system.

The results of the analysis also show the cost effect of controlling SO_2 to different emission levels using FBC technology. The difference in total annual cost is small for the Eastern high sulfur coal and insignificant for the two low sulfur coals. With FBC technology, once a decision is made to control SO_2 emissions to 75 percent or greater, there is fairly small impact in proceeding to more stringent levels, up to 90 percent reduction.

4.3.6 Cost Effectiveness of AFBC SO_2 Control — Unit Cost Basis

The cost of SO_2 control in AFBC is shown in comparison to uncontrolled conventional boiler cost in Table 39 and Figure 38 in terms of \$/kg of SO_2 removed. This parameter accounts for the total annual cost of uncontrolled conventional boilers by subtracting it from the total annual cost of AFBC boilers with SO_2 control (but excluding final particulate control). The balance is divided by the amount of SO_2 removed annually for the set of conditions of concern. As a result, positive values indicate FBC costs are greater, and negative values indicate FBC costs are lower than uncontrolled conventional boilers of the same capacity.

The data illustrate the same trends presented earlier, but give some idea of cost effectiveness. With low sulfur coals, the impact of going to more stringent SO_2 control levels is less than for the case of high sulfur coal. The absolute values are lower, as are the slopes of the curves for low sulfur coal. The linear relationships suggest that even greater levels of SO_2 control (>90 percent) could be achieved without a sharply accelerated cost impact.

It is important to note the impact of sorbent reactivity on control cost (see Table 39). For stringent control using high sulfur coal, the unit costs are shown to vary by about ±\$1.00/kg SO_2 removed for sorbent of high or low reactivity; the variation decreases slightly as boiler capacity decreases,

TABLE 39. COST OF SO_2 CONTROL IN AFBC DOLLARS/KG SULFUR DIOXIDE REMOVED

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW 8.8	 22	 44	 58.6
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	2.04	5.16	4.84	4.13
		LOW	4.2	2.87	6.13	5.72	5.01
		HIGH	2.3	1.11	4.27	3.86	3.15
	I 85%	AVERAGE	2.9	1.63	4.78	4.41	3.71
		LOW	3.8	2.47	5.72	5.29	4.58
		HIGH	2.1	0.89	4.06	3.63	2.92
	M 78.7%	AVERAGE	2.5	1.22	4.37	3.97	3.27
		LOW	3.4	2.05	5.17	4.85	4.14
		HIGH	1.8	0.56	3.74	3.29	2.58
	SIP 56%	AVERAGE	1.0	-0.35	2.85	2.25	1.60
		LOW	1.2	-0.17	3.03	2.43	1.79
		HIGH	0.8	-0.54	2.66	2.07	1.42
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	-0.45	2.75	2.07	1.19
		LOW	3.7	-0.27	2.93	2.25	1.37
		HIGH	2.0	-0.62	2.59	1.91	1.03
	M 75%	AVERAGE	2.2	-0.59	2.61	1.93	1.06
		LOW	3.2	-0.39	2.82	2.13	1.26
		HIGH	1.6	-0.72	2.49	1.81	0.94
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	-2.06	1.69	0.40	-0.31
		LOW	3.6	-1.88	1.87	0.58	-0.14
		HIGH	2.0	-2.20	1.56	0.27	-0.45
	M 75%	AVERAGE	2.2	-2.17	1.56	0.29	-0.42
		LOW	3.2	-1.97	1.78	0.49	-0.23
		HIGH	1.6	-2.29	1.47	0.18	-0.53

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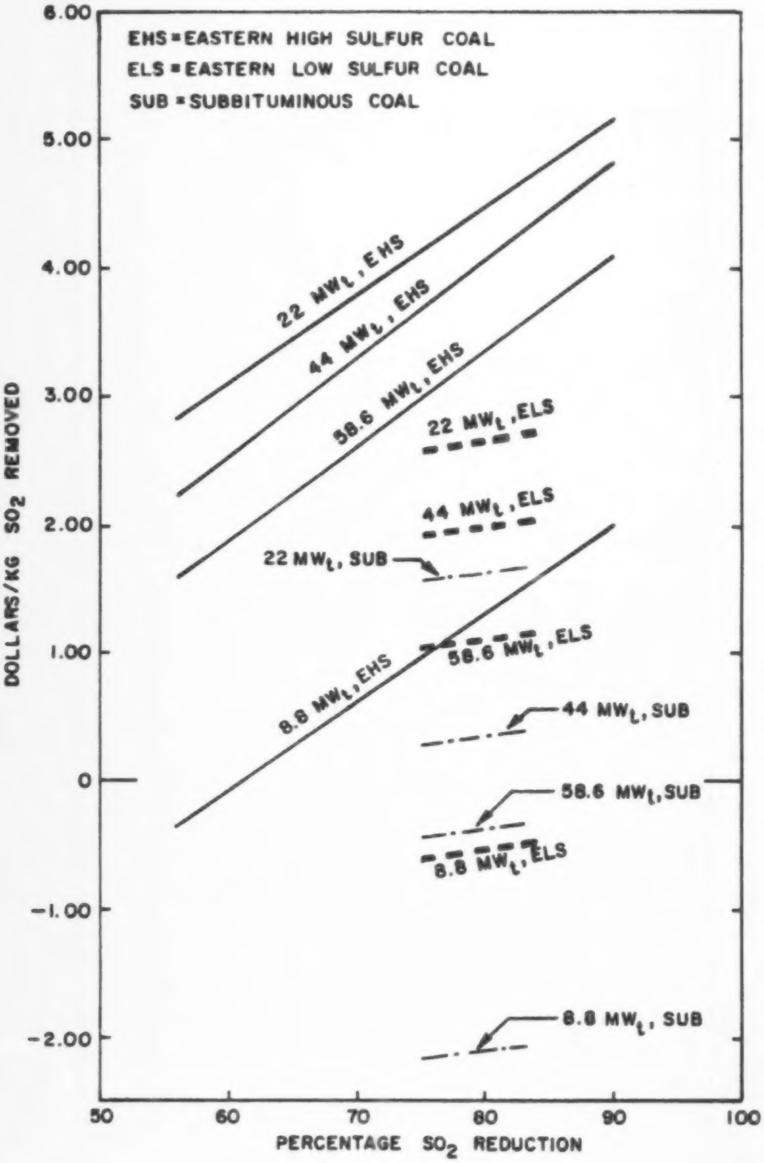

Figure 38. Unit cost of SO_2 control in AFBC boilers with capacity of 8.8 to 58.6 MW_t (30 to 200 × 10^6 Btu/hr).

and decreases greatly as control level decreases. The impact of sorbent reactivity at one particular control level is equivalent to the incremental cost of attaining stringent SO_2 control in comparison to moderate SO_2 control using an average reactivity sorbent. Sorbent reactivity is of somewhat less importance when low sulfur coals are burned.

4.3.7 Comparison of GCA Data with Other Independent Estimates of AFBC Costs

4.3.7.1 Westinghouse Study--

Westinghouse Research and Development is preparing an independent assessment of industrial FBC boiler cost as part of their study, "Effect of SO_2 Emission Requirements on Fluidized-Bed Boilers for Industrial Applications: Preliminary Technical/Economic Assessment."[35] The basis of the cost estimate is intentionally similar to GCA's, as shown in Table 40.

The Westinghouse cost data are presented in Appendix D, Tables D-1 through D-12. Total annual cost in terms of $\$/10^6$ Btu output was estimated by GCA (as shown at the bottom of each table) based on boiler capacity, total annual cost, and boiler efficiency. Annual fixed charge is shown in the same terms, as estimated by GCA from the Westinghouse data. Total turnkey cost was annualized using the same factors used by GCA for capital recovery (0.106) and G&A, taxes, and insurance (0.04).

FBC costs (fixed annual, operating, and total annual) estimated by GCA and Westinghouse (for average sorbent reactivity) are shown comparatively in Figures 39 through 41. The graphs illustrate that Westinghouse estimates a slightly higher total annual cost than GCA for the 8.8 MW_t boiler and significantly lower annual costs for all other boiler capacities. The major difference is in the fixed annual charge for each boiler, since total annual operating costs

TABLE 40. FEATURES OF WESTINGHOUSE COST ESTIMATE FOR INDUSTRIAL FBC BOILERS

Boiler capacities	8.8, 22, 44, 58.6 MW_t
Coal types	Eastern high sulfur, Eastern low sulfur, Subbituminous, as specified in the GCA study.
SO_2 control levels considered	Same as GCA analysis
Cost Basis	Same as GCA analysis, except that costs were based on in-house Westinghouse data and other accessible sources. Boiler vendor quotes were not solicited. Some other differences are: - Limestone purchase cost 25 $/ton - Spent solids disposal cost 8 $/ton
Sorbent types (500 μm average in bed particle size)	I (High Reactivity) - Western 90% CaL II (Medium Reactivity) - Bussen Quarry III (Low Reactivity) - Menlo Quarry

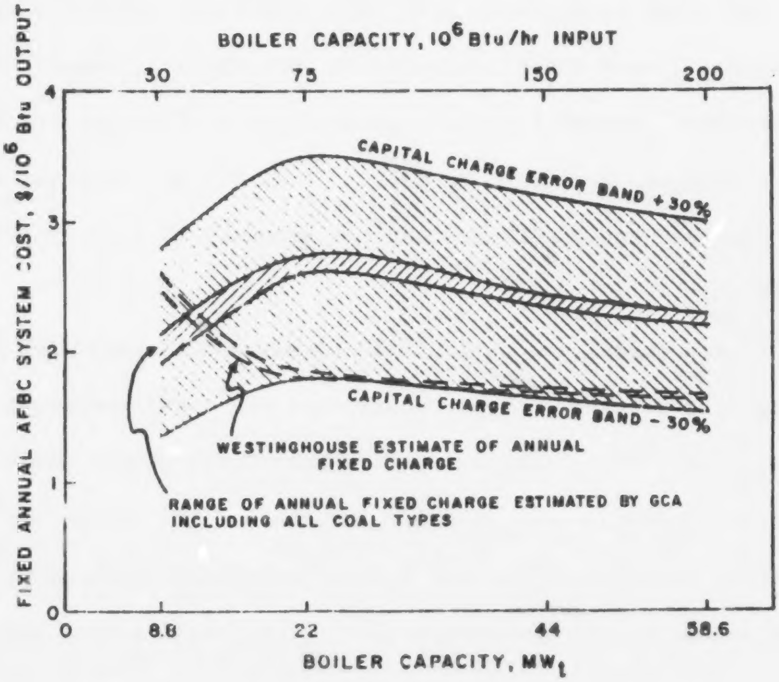

Figure 39. Total fixed annual cost of AFBC with SO_2 control.

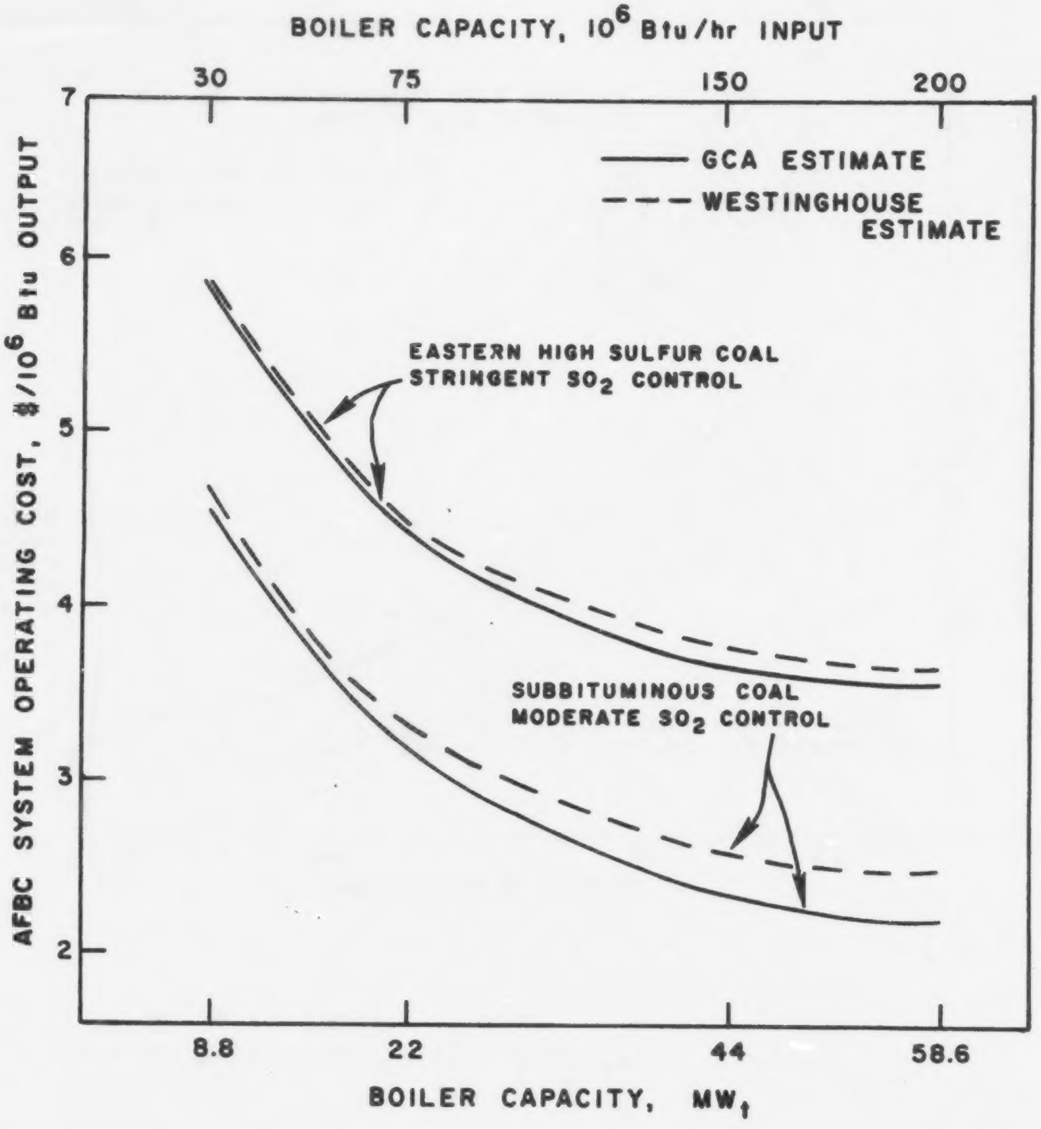

Figure 40. Total operating cost of AFBC with SO_2 control.

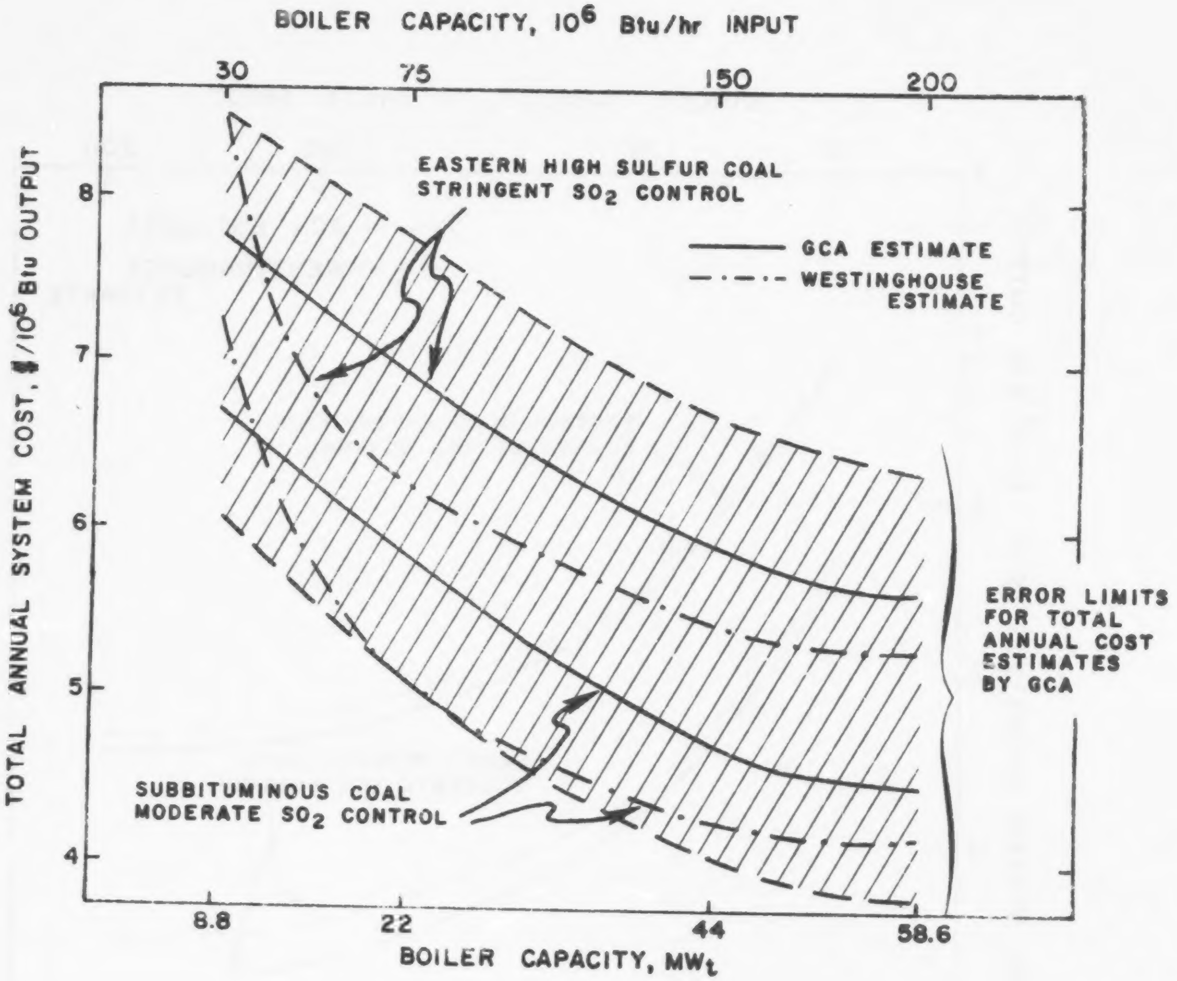

Figure 41. Total annual cost of AFBC with SO_2 control.

(the difference between total annual cost and annual fixed cost) are close to the same for both estimates. It is noted that the capital cost difference is just within the capital charge error band associated with the GCA estimate.

These results poignantly illustrate the possible disparity of two independent cost estimates where the goal of each effort is to maintain a similar cost basis.

It is clear that the absolute values determined in this cost analysis must be used with caution. The differences in the two FBC cost estimates exist because they are budget estimates. They illustrate the accuracy of budget costing procedures and show that the validity of an FBC/conventional boiler cost comparison is very much a function of the source of cost information. Whereas GCA relied on two vendors for AFBC boiler equipment costs (one design for 8.8 MWt, and one design for the other capacities), Westinghouse utilized a similar design for all capacities and based costs on in-house information which they have developed for their continuing studies of fluidized-bed combustion.

When dealing with an emerging technology such as fluidized-bed combustion, the validity of absolute values determined in a budget cost estimate are subject to question. They should not be used for site-specific decisions and should be used cautiosly in any other more general comparison. The merit of this costing procedure lies in the estimation of relative cost differences; i.e., the impact of going to more stringent SO_2 control levels or of using less reactive sorbents.

4.3.7.2 EXXON, and A.G. McKee Studies--

The A.G. McKee[36] estimates are based on the DOE Georgetown University unit in Washington, D.C., the EXXON[37] estimates are for the Gulf Coast, and the GCA estimates are for the midwest.* No attempt was made to adjust costs for location. Some items were adjusted, however, to achieve compatability with the assumptions in this study, but care was taken to maintain the integrity of the other estimates. Appendix B presents the basis of other cost studies and describes the adjustments made by GCA.

Table 41 presents a summary of annual costs in terms of $\$/10^6$ Btu output for AFBC burning Eastern high sulfur coal. The estimates represent an SO_2 removal efficiency of 85 percent to be comparable with a limit of 516 ng/J (1.2 $lb/10^6$ Btu) specified by EXXON and a Ca/S ratio of 3 specified by A.G. McKee. Figures 42 and 43 graphically illustrate the cost data.

The EXXON values (updated from 1975) are in agreement with the GCA estimates for total annual cost and annual fixed charges assuming that interpolation of GCA data is valid. The A.G. McKee estimates are significantly lower than GCA or EXXON, probably for two reasons. First, the Georgetown unit is being installed (startup began July, 1979) as an additional boiler at an existing facility so that several equipment items normally required at a "grass roots" location are not necessary. This would include the steam circulation system, and boiler feedwater treatment. Coal and solid waste handling are necessary, however, because the two existing boilers are natural gas/oil-fired units. Second, since the unit is currently being erected, contingencies that must be added to budget estimates may not be applicable for the Georgetown unit. It is not possible to conclude whether the McKee cost data validate the

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*Based on the assigned groundrules of the overall EPA Industrial Boiler Study.

TABLE 41. AFBC BOILER COST WITH 85 PERCENT SO_2 REMOVAL

Source	Annual cost, $/$10^6$ Btu output				
	Plant size - MW_t				
	8.8	22	37	44	58.6
GCA (controlled AFBC)	7.62	6.84		5.78	5.56
EXXON (controlled AFBC)	-	-	6.14	-	-
A.G. McKee (conventional with no SO_2 control)	-	-	4.34*	-	-
A.G. McKee (controlled AFBC)	-	-	4.71	-	-

*Annual fixed charge for this estimate is $2.30/$10^6$ Btu output.

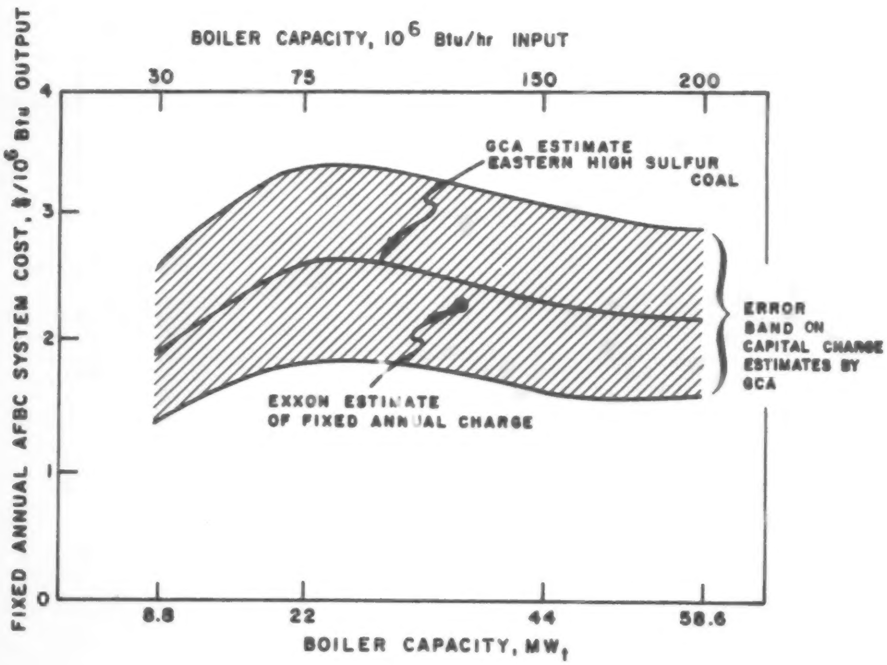

Figure 42. Comparison of fixed annual cost estimates.

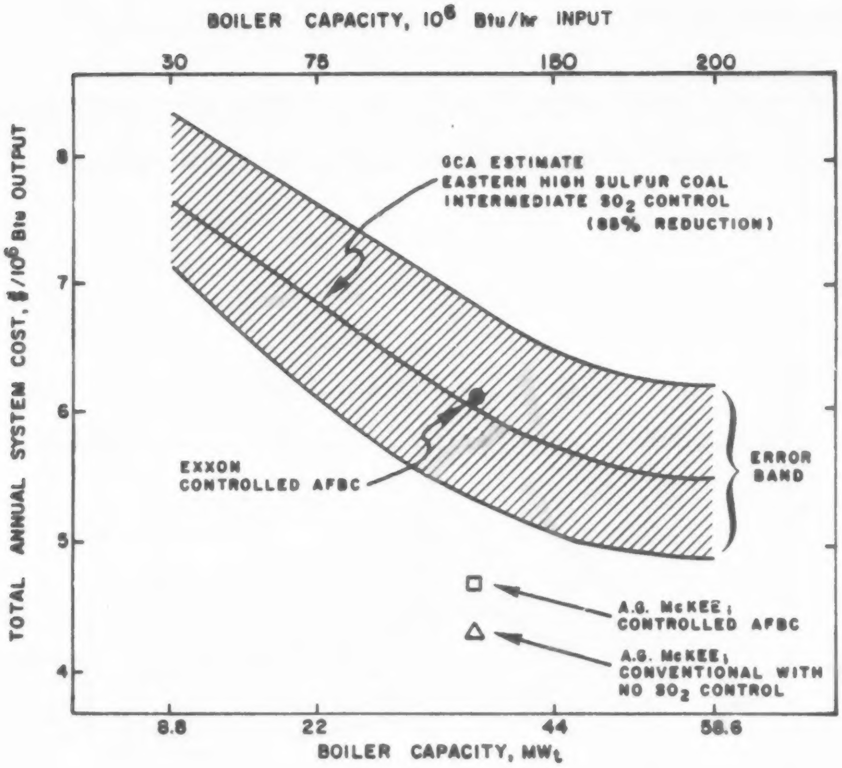

Figure 43. Comparison of total annual cost estimates.

GCA estimates since the impact of the two factors mentioned above can not be quantified. However, the EXXON estimates do support the GCA cost values.

The difference in cost between the AFBC boiler with SO_2 control and conventional boiler without SO_2 control, as estimated by A.G. McKee is small, amounting to 8 percent. This difference is slightly less than that noted in the earlier comparison of GCA and PEDCo derived costs for controlled AFBC and uncontrolled conventional systems, respectively.

4.3.8 Sensitivity Analysis - Cost

An analysis of the cost sensitivity of AFBC (incorporating "best system" design/operating conditions) to variations in operating parameters, raw material costs, and capital costs was performed. The results are reported as dollars per million Btu output ($\$/10^6$ Btu) and, where appropriate, as dollars per kilogram sulfur dioxide removed ($\$/kg\ SO_2$ removed). This analysis required definition of a baseline set of conditions which is presented in Appendix C, Table C-3. These conditions are representative of high sulfur coal combustion, with an average sorbent and stringent SO_2 control. The various operating conditions investigated were: effect of heat recovery, plant load factor, excess air, combustion efficiency, calcium-to-sulfur ratio, moisture removal requirements, sulfur capture, and coal sulfur content. Materials and capital cost effects which were investigated were coal cost, limestone cost, residue disposal cost, and variation in capital expenditure due to design changes. The ranges investigated are also listed in Table C-3.

Of the parameters investigated, seven exhibited linear relationships, three exhibited nonlinear relationships, and two had an insignificant effect on cost. The seven linear variables are:

- coal cost
- limestone cost
- residue disposal cost
- capital cost
- Ca/S ratio
- coal drying
- coal sulfur

The predictive equations are presented in Table 42.

The linear variables are discussed in three groups. Coal cost, limestone cost and residue disposal cost are presented together as purchase or disposal costs. Capital cost is discussed separately. Ca/S ratio, drying requirements and coal sulfur content are discussed as operating variables.

The nonlinear variables are:

- combustion efficiency
- excess air
- plant load factor

Combustion efficiency and excess air are discussed together because their effects are interactive (changing one forces variation in the other). Plant load factor is discussed separately because this is a function only of steam demand from the industrial user.

Neither of the other two variables investigated (heat recovery and sulfur capture) had a significant effect on cost. Spent solids heat recovery, a feature incorporated in some designs, decreased costs by 0.07 $\$/10^6$ Btu as heat recovery varied from 0 to 100 percent. Sulfur capture is calculated under the assumption that the Ca/S ratio remains constant, and the only variation is sorbent reactivity. When sulfur capture changes from 70 to 90 percent, cost is reduced by only 1¢/10^6 Btu. Neither variable is significant for industrial considerations.

TABLE 42. GENERAL EQUATIONS RELATING COAL COST, LIMESTONE COST, RESIDUE DISPOSAL COST, CAPITAL COST, Ca/S RATIO, DRYING, AND COAL SULFUR TO $\$/10^6$ BTU

Parameter	Method of firing	$\$/10^6$ BTU for plant size of			
		8.8 MW_t	22 MW_t	44 MW_t	58.6 MW_t
Coal Cost	Uncontrolled Conventional	0.0562C + 6.49	0.0558C + 4.68	0.0554C + 3.88	0.0532C + 3.71
	AFBC	0.0548C + 6.90	0.0542C + 6.16	0.0540C + 5.08	0.0538C + 4.86
Limestone Cost	Uncontrolled Conventional	—	—	—	—
	AFBC	0.023L + 7.60	0.023L + 6.65	0.023L + 5.77	0.023L + 5.55
Residue Disposal Cost	Uncontrolled Conventional	0.0046R + 7.22	0.0043R + 5.60	0.0020R + 4.69	0.011R + 4.51
	AFBC	0.022R + 6.94	0.0022R + 6.19	0.022R + 5.11	0.022R + 4.89
Capital Cost	Uncontrolled Conventional	—	—	—	—
	AFBC	0.0186W + 6.02	0.0258W + 4.68	0.0228W + 3.79	0.0212W + 3.73
Ca/S Ratio	Uncontrolled Conventional	—	—	—	—
	AFBC	0.379Ca/S + 6.98	0.373Ca/S + 6.22	0.355Ca/S + 5.11	0.351Ca/S + 4.84
Drying Requirement	Uncontrolled Conventional	0.0203M + 7.35	0.018M + 5.72	0.0163M + 4.73	0.0190M + 4.49
	AFBC	0.0263M + 7.75	0.0247M + 6.99	0.0220M + 5.91	0.0215M + 5.70
Coal Sulfur	Uncontrolled Conventional	0.0044S + 7.38	0.0033S + 5.75	0.0033S + 4.76	0.0022S + 4.56
	AFBC	0.368S + 6.50	0.368S + 5.77	0.359S + 4.69	0.356S + 4.48

M = Percent Moisture Removed

C = Coal Cost

W = Percent of Original Estimate

Ca/S = Calcium-to-Sulfur Ratio

L = Limestone Cost

R = Residue Disposal Coast

S = Coal Sulfur Content

4.3.8.1 Material Cost Variation--

Coal cost, limestone cost, and residue disposal cost are all site-specific costs. No adjustment for waste reuse (such as road bed filler or agricultural applications) was attempted because these uses are not only site-specific, but also seasonal.

The linear equations shown in Table 47 can be used to determine the cost of steam for any hypothetical site under investigation. Consider a site with coal costing \$22/ton, limestone at \$14.90/ton and residue disposal at \$31/ton. The base costs (see Tables C-3 and C-4) are respectively: coal - \$17/ton, limestone - \$8/ton, and residue disposal - \$40/ton. Using coal cost as the standard equation from Table 47, the cost in dollars per million Btu output is approximated as follows:

- Conventional Spreader Stoker

$$\begin{aligned}\$/10^6 \text{ Btu} &= 0.0554C + 3.88 + 0.0020\,(R-40)\\ &= 0.0554\,(22) + 3.88 + 0.0020\,(31-40)\\ &= 5.08\end{aligned}$$

- AFBC

$$\begin{aligned}\$/10^6 \text{ Btu} &= 0.0540C + 5.08 + 0.023\,(L-8) + 0.022\,(R-40)\\ &= 0.054\,(22) + 5.08 + 0.023\,(14.9-8) + 0.022\,(31-40)\\ &= 6.23\end{aligned}$$

where
C = coal cost
L = limestone cost
R = residue disposal cost

The calculated differential of \$1.15 indicates that a controlled AFBC boiler produces steam at a cost 23 percent higher than an uncontrolled conventional boiler under these hypothetical conditions. The significance of this difference is questionable when one considers that cost estimate accuracy limits are specified as ±30 percent.

4.3.8.2 Capital Investment Variation--

The linear relationship for capital cost variation in Table 47 predicts the cost of design variations specifically affecting the AFBC cost estimates. The capital cost variation analysis should not be confused with the aforementioned estimated accuracy limits of ±30 percent.

Cost estimate accuracy limits pertain to errors in overall cost estimates. The capital cost variation analysis is designed to determine the effect on output cost when design changes (such as in-bed fuel feeding or deeper beds) increase the anticipated capital cost. Because the focus of this report is comparison of AFBC steam costs with conventional steam costs, only the capital cost of those items unique to FBC were varied. Items common to both systems (such as coal handling equipment) and items unique to conventional firing (such as the conventional firebox) are held constant.

An example of the use for which this analysis is intended is the cost effect of replacing stoker feed AFBC with underbed feed AFBC. If preliminary cost analysis indicates in-bed feed adds 20 percent to the system capital investment, the cost of steam increases by $0.40/10^6 Btu for the large boiler (58.6 MW_t).

4.3.8.3 Operating Variations--

Sorbent requirements at a specific control level are a function of system design, sorbent reactivity, coal sulfur, and sorbent particle size. The coal sulfur effect in terms of \$/10^6 Btu output is linear and the equations are presented in Table 47. Rigid relationships linking the other three parameters (system design, sorbent reactivity, and sorbent particle size) to cost are not well defined.

Figure 44 illustrates the effect of coal sulfur content on cost in terms of \$/kg SO_2 removed. The nonlinear curves result because the cost of conventional boilers is subtracted from the total AFBC cost and the balance is divided by the annual amount of SO_2 removed. For low sulfur coal, the cost per unit sulfur dioxide removed is quite dependent on coal sulfur content. Above 4 percent sulfur, the relationship is nearly linear.

Changes in system design to alter "commercially offered" systems to "best systems" as defined in this study are increased gas phase residence time and reductions in sorbent particle size. These changes reduce sorbent requirements by enhancing the gas/solid reaction.

The linear equations in Table 47 can predict cost effects of reduced Ca/S requirements. For instance, if a commercial design requires a Ca/S ratio of 3.5 and the "best system" would require a Ca/S ratio of only 2.5, the cost reduction is \$0.35 to \$0.37/10^6 Btu depending upon boiler size (assuming no capital cost changes). Coal drying (removal of surface moisture) is a requirement for AFBC only if an underbed feed design is necessary for maintenance of low emissions. From the equations in Table 47, every incremental reduction of 5 percent moisture increases cost by \$0.10/10^6 Btu output.

4.3.8.4 Nonlinear Effects in Cost Estimates--

Three of the variables investigated are nonlinear in cost of heat produced. These are combustion efficiency, excess air, and plant load factor.

Figures 45, and 46 depict the interrelationship between cost and: (1) combustion efficiency; or (2) excess air. The cost of conventional firing under the standard design assumptions is included at the reference conditions noted in Table C-2. Although the relationship in both cases is nonlinear, the deviation from linearity is minor. Assuming combustion efficiency drops from

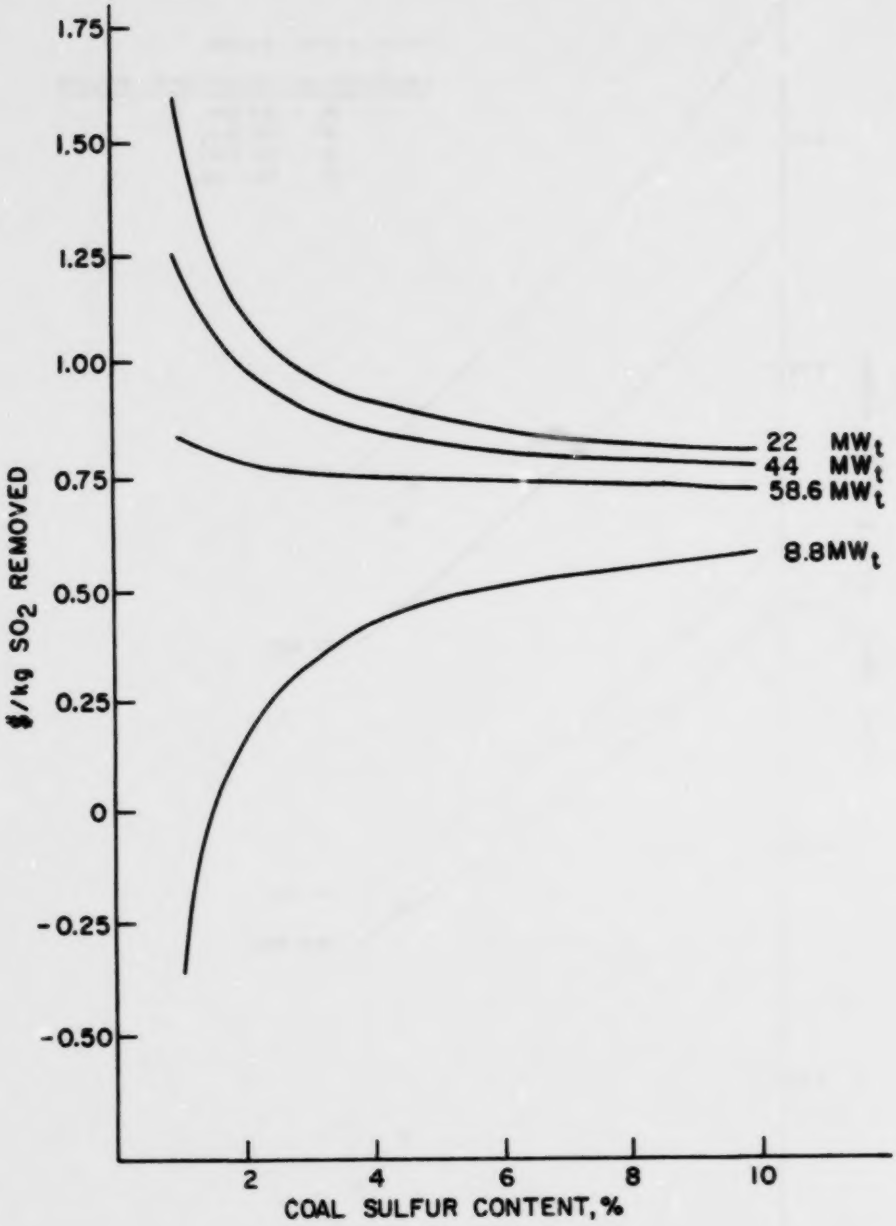

Figure 44. Cost of SO_2 control in AFBC ($/kg SO_2 removed) versus coal sulfur content.

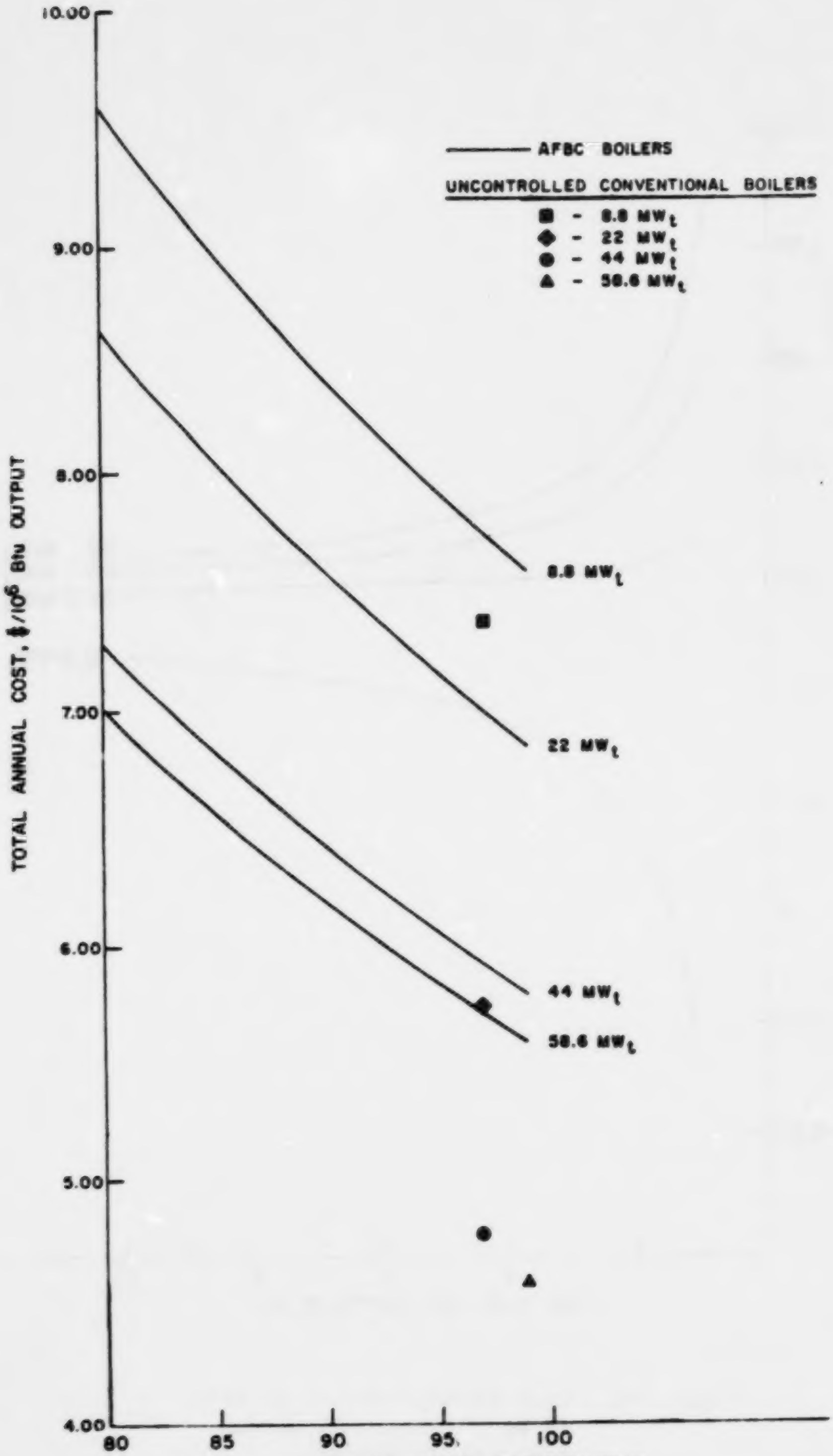

Figure 45. Cost of AFBC with SO_2 control versus combustion efficiency.

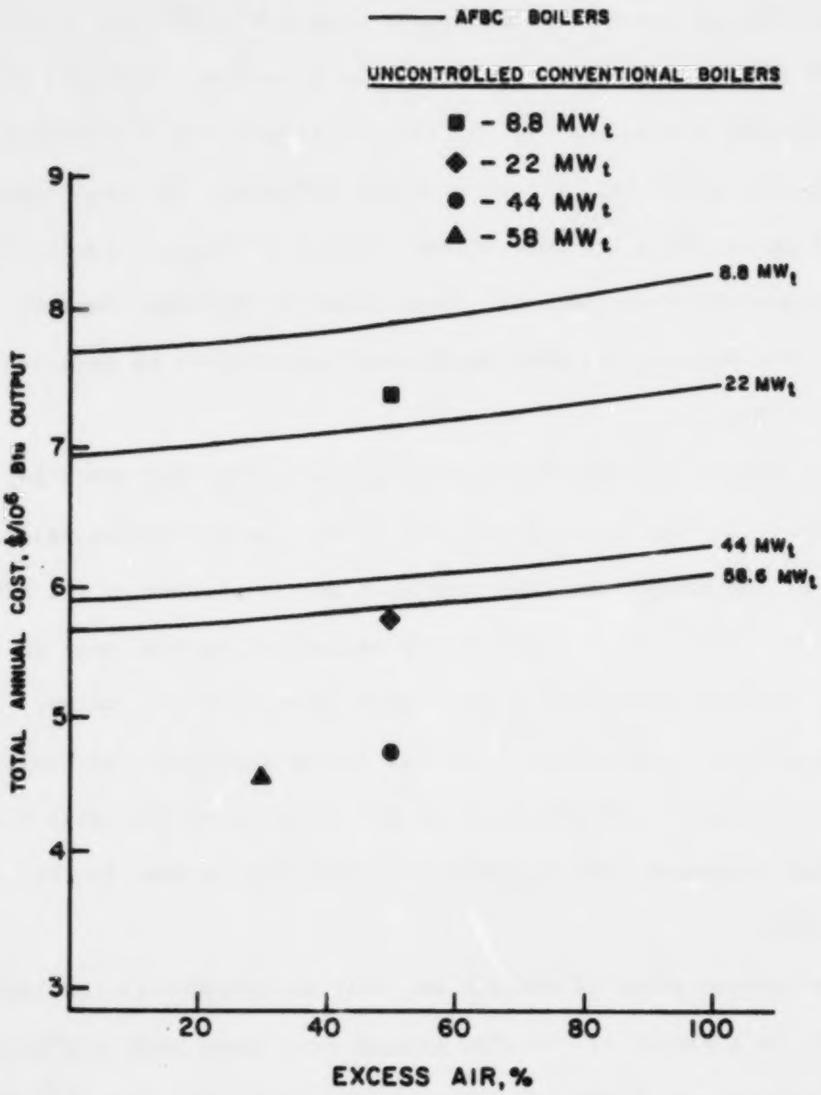

Figure 46. Cost of AFBC with SO_2 control versus excess air.

95 percent down to 90 percent, the projected cost increases from 5.86 $\$/10^6$ Btu to 6.19 $\$/10^6$ Btu for the 58.6 MW_t unit. Based on linear regression analysis, the projected cost increases from 5.87 $\$/10^6$ Btu to 6.24 $\$/10^6$ Btu.

The effect of load factor on cost is presented in Figure 47 and 48 in terms of $\$/10^6$ Btu output (22 MW_t case only) and \$/kg SO_2 removed. The ordinate of Figure 47, $\$/10^6$ Btu, illustrates the difference in cost between controlled AFBC and an uncontrolled conventional boiler of 22 MW_t capacity. In Figure 48, the ordinate, \$/kg SO_2 removed, is obtained by dividing the cost difference in $\$/10^6$ Btu between the AFBC and conventional boiler by emissions in terms of kg $SO_2/10^6$ Btu.

From Figure 55, the effect of both the capital and operating cost components is evident. At low load factor; e.g., 0.40, annualized capital comprises 27 percent of the conventionally-fired cost and 30 percent of the AFBC cost. At 100 percent load, conventional-firing annualized capital cost is 23 percent. Similar analysis of the other three capacities, 8.8 MW_t, 44 MW_t, and 58.6 MW_t, produces similar trends; i.e., as load factor increases the capital component to cost decreases. Additionally, as the fraction of the total cost attributed to capital decreases, the dependence of \$/kg SO_2 on load factor decreases (Figure 48).

The inverse slope of the 8.8 MW_t unit as compared to the larger units in Figure 48 is a result of the AFBC capital cost comprising a significantly smaller proportion of the total annual AFBC cost than does the capital cost of the conventional unit (see Figure 47). As a result, when load factor increases, AFBC costs increase more rapidly than conventional costs because incremental operating costs are higher for AFBC than for conventional uncontrolled systems.

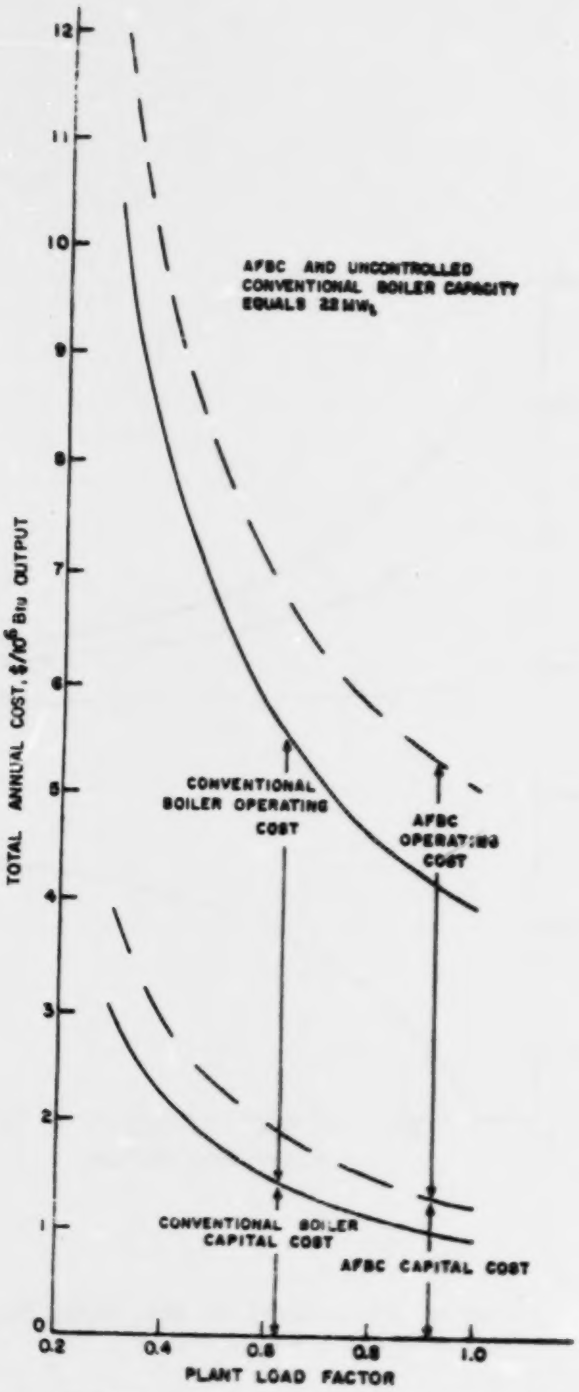

Figure 47. Cost of AFBC at a Capacity of 22 MW$_t$ with SO_2 control versus plant load factor.

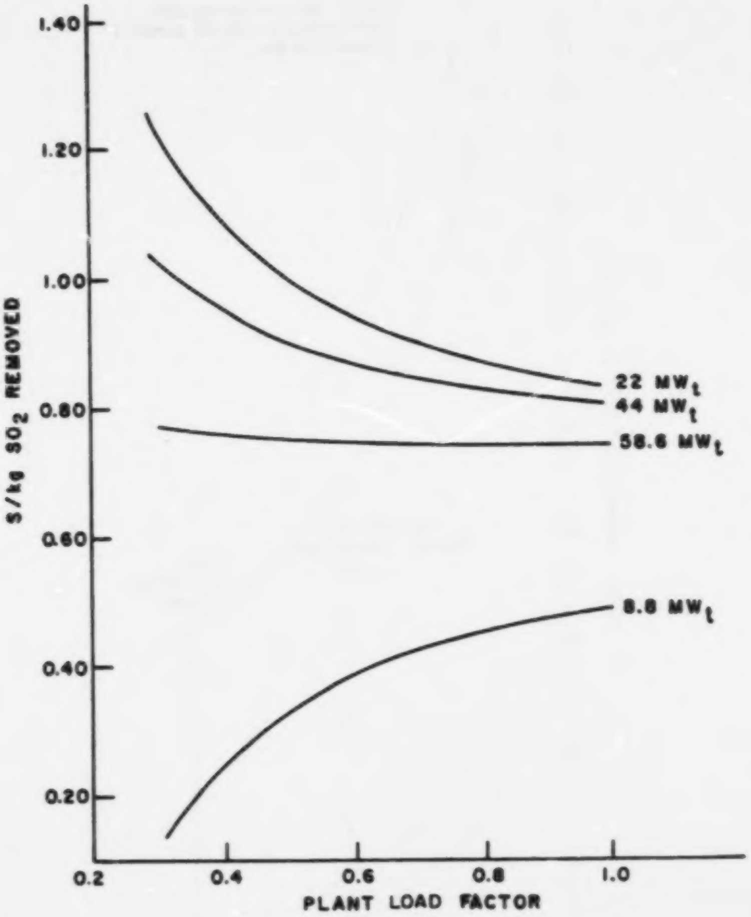

Figure 48. Cost of SO_2 control in AFBC ($/kg SO_2 removed) versus plant load factor.

The parameters investigated, the range investigated, and the resultant cost ranges are presented in Table 43. This table may be used to assign qualitative rankings to the variables with regard to their effect on cost. However, even though the range investigated is within the limits one could expect to encounter, the entire range would not be expected to occur at one site. For example, coal cost can easily range from \$10/ton up to \$60/ton, but the limits for a specific site or specific coal would not ordinarily be this wide.

Considering this qualification, the major variables are load factor, coal cost, combustion efficiency, Ca/S ratio, and coal sulfur. Intermediate variables are drying requirement, capital cost, excess air, limestone cost, and residue disposal cost. Relatively insignificant variables are heat recovery and sulfur capture. (Sulfur capture is insignificant in this analysis because Ca/S ratio was held constatn at 3.5). This prioritization of variables provides an insight into the significance of each variable investigated. However, the significance of each factor for a specific site depends on the price range for the locale.

4.4 COST OF BEST SYSTEM PARTICULATE CONTROL FOR COAL-FIRED AFBC INDUSTRIAL BOILERS

4.4.1 Attempt to Isolate Particulate Control Costs from SO_2 Control Costs

Final particulate control cost is reported in this section for AFBC boilers with and without SO_2 control. Although we do not anticipate that coal-fired AFBC industrial boilers without SO_2 control will be used to a significant extent, the analysis of cost of particulate control applied to such systems is presented for the sake of completeness.

Definition of the cost of AFBC without SO_2 control but with particulate control is difficult since AFBC is inherently a combined energy production/SO_2 control technology (see Subsection 4.1.4.1). It was roughly estimated for the

TABLE 43. COST SENSITIVITY ANALYSIS - AFBC

Parameter	Range studied	Cost range, $/10^6 Btu output			
		8.8 MWt	22 MWt	44 MWt	58.6 MWt
Drying requirement	0 - 30 percent	7.75 - 8.50	6.99 - 7.69	5.91 - 5.86	5.70 - 6.31
Heat recovery	250 - 1480°F	7.69 - 7.79	6.95 - 7.03	5.88 - 5.95	5.66 - 5.74
Load factor	0.30 - 1.00	13.49 - 5.50	12.01 - 5.04	9.85 - 4.38	9.42 - 4.25
Coal cost	10.00 - 60.00 $/ton	7.42 - 10.00	6.67 - 9.24	5.59 - 8.15	5.37 - 7.92
Capital cost	0.60 - 1.40	7.08 - 8.49	6.06 - 8.01	5.08 - 6.80	4.93 - 6.53
Excess air	0 - 100 percent	7.68 - 8.27	6.95 - 7.47	5.88 - 6.31	5.66 - 6.08
Combustion efficiency	80 - 99 percent	9.56 - 7.62	8.61 - 6.88	7.28 - 5.82	7.00 - 5.61
Ca/S ratio	1 - 10	6.96 - 10.15	6.24 - 9.38	5.17 - 8.16	4.95 - 7.91
Sulfur capture	70 - 100 percent	7.78 - 7.79	7.02 - 7.03	5.93 - 5.95	5.72 - 5.73
Limestone cost	5 - 35 $/ton	7.71 - 8.38	6.96 - 7.63	5.89 - 6.54	5.66 - 6.32
Residue disposal cost	5 - 40 $/ton	7.04 - 7.78	6.29 - 7.03	5.21 - 5.94	5.00 - 5.72
Coal sulfur content	1 - 10 percent	6.85 - 9.98	6.12 - 9.25	5.03 - 8.09	4.84 - 7.86

purpose of this section by omitting limestone handling and purchase costs. Spent solids handling and disposal costs were modified to allow only for withdrawal of bed bottom ash. Total auxiliary power was reduced by 15 percent to estimate electricity requirements when SO_2 control is not practiced.

Available data on particulate emissions and control efficiency for AFBC are limited relative to data for conventional systems; therefore, differences between FBC and conventional control system costs cannot be quantified. Factors that could cause differences in equipment design, applicability, or cost are discussed in Sections 2.0 and 3.0.

Particulate control device costs developed in the Particulate Control ITAR[38] are considered to be representative for application to FBC boilers, accounting for the error bounds of the cost estimates used in this study, which are estimated to be ±40 percent for the combined AFBC boiler and particulate control devices.

To determine the cost of particulate control for AFBC boilers employing SO_2 control with limestone, it was assumed that particulate control device costs for FBC are the same as for conventional boilers burning low sulfur coal. For particulate control cost in AFBC boilers not controlling SO_2, costs were assumed to vary depending on specific coal type.

4.4.1.1 Inlet Particle Loadings--

The discussion in Sections 2.0 and 3.0 pointed out that data are limited on particulate loadings from atmospheric fluid-bed units. Considering the existing data base, it is estimated that uncontrolled particulate emissions (i.e., loadings downstream of the primary cyclone) will range between 215 to 2150 ng/J (0.5 to 5.0 $lb/10^6$ Btu) in systems operating under "best" conditions for SO_2 control.* Variation within the range will depend on primary cyclone efficiency, the level of SO_2 control, in-bed particle size distribution, coal ash, freeboard height, the effect of baffling by the convection pass of heat transfer tubes, the extent of recycling, and other considerations. Whether uncontrolled particle loadings fall below this range if SO_2 control is not practiced is unclear because of the number of influential factors in addition to sorbent loading and sorbent particle size. In PER testing of the FBM[39] (see Section 7.0, Table 84), uncontrolled particulate emissions were measured in the range of 430 to 730 ng/J (1.0 to 1.7 $lb/10^6$ Btu) when sorbent was added for SO_2 control. Burning the same coal without sorbent addition, particle emissions were measured to range between 301 to 559 ng/J (0.7 to 1.3 $lb/10^6$ Btu). This reduction is significant but is still above the minimum specified earlier. The fact that grain loading was not reduced even further is of interest because the sorbent used for SO_2 reduction was fed at a top size of 44 μm. This implies that other factors are influential in determining uncontrolled particulate emissions, and that estimation of particle loadings on a general basis when SO_2 control is not practiced, cannot be done without more thorough data. Other comparative data for a single unit are not available.

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*Expanded bed depth = 1.2 m (4 ft); superficial gas velocity = 1.8 m/sec (6 ft/sec); average in-bed sorbent size = 500 μm.

Therefore, to estimate the cost of particulate control for AFBC systems with and without SO_2 control, we have assumed a common uncontrolled particle emission range between 215 to 2150 ng/J (0.5 to 5.0 $lb/10^6$ Btu). This assumption could be a source of error for the estimates of ESP cost since ESP design is a strong function of particle loading and particle chemistry. It should not be a source of error for fabric filters or multitube cyclones since the cost of these devices is more strongly related to flue gas volume.

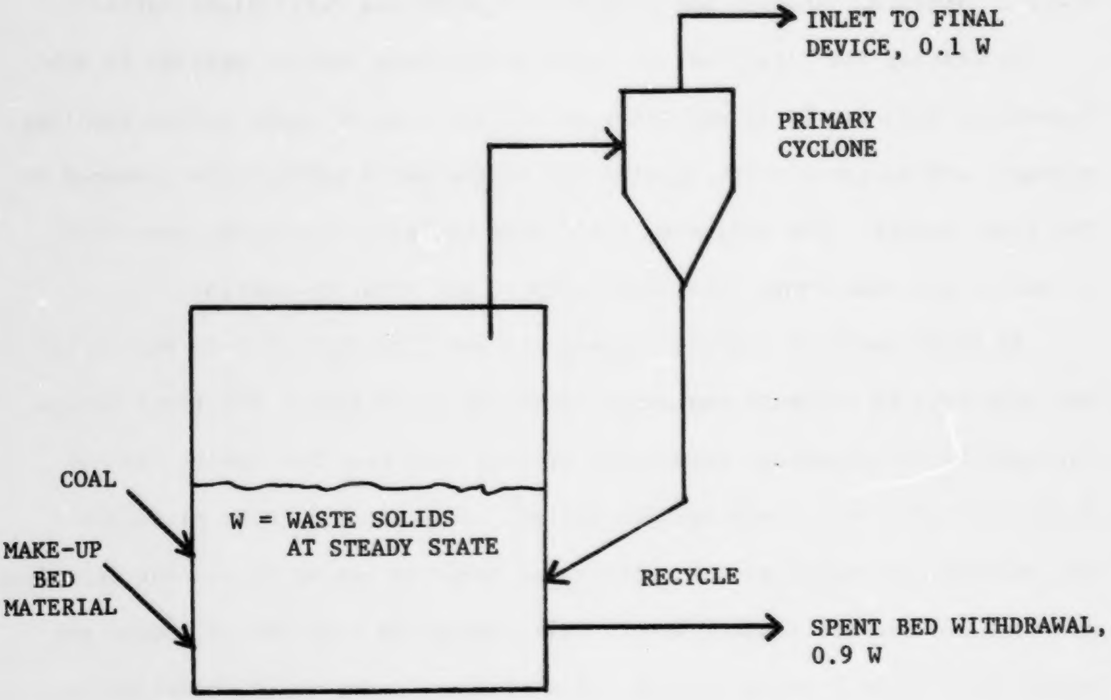
4.4.1.2 Handling, Storage and Disposal of Collected Particulate Matter--

To develop the total cost of final particulate control applied to AFBC industrial boilers, it is necessary to add the cost of waste solids handling, storage, and disposal which results due to the added particulate captured by the final device. The following discussion explains how these costs were estimated for each case; i.e., AFBC without and with SO_2 control.

An inlet particle loading between 215 and 2150 ng/J (0.5 to 5.0 $lb/10^6$ Btu) was used to estimate the range of solids collected by the final device for each boiler capacity, regardless of coal type when SO_2 control is not practiced. For this analysis, 100 percent collection of inlet particulate was assumed. Although actual capture can range as low as 50 percent depending on inlet loading and control level, this assumption does not introduce any significant error because the cost of additional spent solids handling is generally less than 2 percent of the total cost of AFBC with particulate control.

A factor of \$40/ton was applied to estimate the cost of additional spent solids disposal. A unit cost factor ranging between \$8.60 to \$12.20 ft^3 of storage capacity was used to estimate the cost of added handling and storage (see Subsection 4.3.1). Appropriate factors were applied to account for direct and indirect installation of handling and storage equipment (see Subsection 4.2.1).

To estimate approximate inlet loadings when SO_2 control is practiced, the system was modeled as follows:


INLET TO FINAL DEVICE, 0.1 W
PRIMARY CYCLONE
COAL
W = WASTE SOLIDS AT STEADY STATE
MAKE-UP BED MATERIAL
RECYCLE
SPENT BED WITHDRAWAL, 0.9 W

where W is the sum of:

- Coal ash
- Unburned carbon
- Limestone inerts
- Uncalcined limestone
- Unused calcium oxide
- $CaSO_4$ produced

The rate, W, for each combination of boiler capacity, coal type, and SO_2 control level is shown in Section 6, Table 80. The ratio of 0.1 W was selected to calculate the inlet final particle loading because the resultant loadings fall within the range of 365 to 1850 ng/J (0.85 to 4.3 lb/10^6 Btu) which approximates the experimentally documented range.

The cost of incremental needs for spent solids handling, storage, and disposal were then calculated by assuming that the final device operated at

100 percent efficiency. For the case of combined particulate and SO_2 control, the low incremental cost is based on the minimum waste production rate, W, for each boiler capacity, which occurs under a moderate SO_2 control level, burning Eastern low sulfur coal with a high reactivity sorbent. The high point occurs under stringent SO_2 control, burning Eastern high sulfur coal with a low reactivity sorbent. These assumptions allowed for ease of computation, since the applicable costs could be directly proportioned from the cost tables which appear in Appendix C for spent solids handling and storage and spent solids disposal cost when SO_2 control is practiced.

4.4.2 Cost of Particulate Control for AFBC Boilers - Excluding Influence of SO_2

The estimated cost of particulate control for AFBC boilers operating without SO_2 control is shown in Table 44, based on vendor quotes for application of the devices to conventional boilers.[40] One exception is the multitube cyclone cost for the 40 MW_t boiler, which resulted from a study conducted by IGCI for the Environmental Protection Agency.[41]

Only those control device/control level combinations are shown which were considered as potential "best systems" in Section 3.0. Total costs are presented in comparison to the cost of uncontrolled conventional boilers as a percentage. This value was calculated as:

$$\begin{array}{c}\text{Percentage}\\ \text{increase in}\\ \text{cost oven}\\ \text{uncontrolled}\\ \text{conventional}\\ \text{boiler}\end{array} = \frac{\left[\begin{array}{c}\text{Total annual cost of AFBC without}\\ \text{SO}_2\text{ control but with}\\ \text{particulate control}\end{array}\right] - \left[\begin{array}{c}\text{Total annual cost}\\ \text{of uncontrolled}\\ \text{conventional boilers}\end{array}\right]}{\left[\begin{array}{c}\text{Total annual cost of}\\ \text{uncontrolled conventional}\\ \text{boiler}\end{array}\right]} \times 100$$

TABLE 44. ESTIMATED COST OF FINAL PARTICULATE CONTROL FOR AFBC BOILERS – EXCLUDING SO_2 CONTROL

Boiler capacity (MW_t)	Control device	Particulate control level	Coal type	Annual cost of final particulate control device	Annual cost of incremental spent solids handling storage and disposal	Approximate annual cost of AFBC without SO_2 control	Approximate total cost of AFBC with particulate control but no SO_2 control	Percentage increase in cost over uncontrolled conventional boilers	$/kg particulate removed
8.8	FF	S, I, M, SIP	All	51,000	1,800 - 17,700	847,000 - 887,000	899,800 - 955,700	(-3.7) - 3.8	(-1.0) - 0.1
22	FF	S, I, M, SIP	All	86,000	4,400 - 43,800	1,860,000 - 2,018,000	1,950,400 - 2,148,000	6.2 - 21.8	1.3 - 0.4
44	FF	S, I, M, SIP	All	147,000	8,900 - 89,100	3,001,000 - 3,333,000	3,156,900 - 3,569,100	2.2 - 17.9	0.4 - 0.3
58.6	FF	S, I, M, SIP	All	181,000	11,900 - 118,500	3,804,000 - 4,276,000	3,996,900 - 4,575,600	(-3.6) - 14.4	(-0.6) - 0.2
8.8	MC	M, SIP	All	10,000	1,800 - 17,700	847,000 - 887,000	858,800 - 914,700	(-8.0) - (-0.7)	(-2.1) - 0
40	MC	M, SIP	All	48,000	8,100 - 81,000	2,756,000 - 3,055,000	2,812,100 - 3,184,000		
44	MC	M, SIP	All	26,000	8,900 - 89,100	3,001,000 - 3,333,000	3,035,900 - 3,448,100	(-1.8) - 13.9	-0.3 - 0.2
8.8	ESP	S	EHS	25,000	1,800 - 17,700	874,000	900,800 - 916,700	(-2.8) - (-1.1)	(-0.7) - 0
8.8	ESP	S	ELS	62,000	1,800 - 17,700	887,000	950,800 - 966,700	3.2 - 5.0	0.8 - 0.1
8.8	ESP	S	SUB	75,000	1,800 - 17,700	847,000	923,800 - 939,700	(-1.1) - 0.6	(-0.3) - 0
8.8	ESP & MC	I	EHS	30,000	1,800 - 17,700	874,000	905,800 - 921,700	(-2.3) - (-0.6)	(-0.6) - 0
8.8	ESP & MC	I	ELS	56,000	1,800 - 17,700	887,000	944,800 - 960,700	2.6 - 4.3	0.7 - 0.1
8.8	ESP & MC	I	SUB	63,000	1,800 - 17,700	847,000	911,800 - 927,700	(-2.4) - (-0.7)	(-0.6) - 0
22	ESP	S	EHS	55,000	4,400 - 43,800	1,962,000	2,021,400 - 2,060,800	10.7 - 12.9	2.2 - 0.3
22	ESP	S	ELS	127,000	4,400 - 43,800	2,018,000	2,149,400 - 2,188,800	17.1 - 19.2	3.4 - 0.4
22	ESP	S	SUB	147,000	4,400 - 43,800	1,860,000	2,011,400 - 2,050,800	14.1 - 16.3	2.7 - 0.3
44	ESP	S	EHS	114,000	8,900 - 89,100	3,222,000	3,344,900 - 3,425,100	9.9 - 12.5	1.7 - 0.2
44	ESP	S	ELS	204,000	8,900 - 89,100	3,333,000	3,545,900 - 3,626,100	14.7 - 17.3	2.6 - 0.3
44	ESP	S	SUB	208,000	8,900 - 89,100	3,001,000	3,217,900 - 3,298,100	6.3 - 8.9	1.1 - 0.2
44	ESP & MC	I	EHS	121,000	8,900 - 89,100	3,222,000	3,351,900 - 3,432,100	10.1 - 12.7	1.7 - 0.2
44	ESP & MC	I	ELS	213,000	8,900 - 89,100	3,333,000	3,554,900 - 3,635,100	15.0 - 17.6	2.6 - 0.3
44	ESP & MC	I	SUB	226,000	8,900 - 89,100	3,001,000	3,235,900 - 3,316,100	6.9 - 9.5	1.2 - 0.2
58.6	ESP	S	EHS	128,000	11,900 - 118,500	4,113,000	4,252,900 - 4,359,500	5.4 - 8.0	0.9 - 0.1
58.6	ESP	S	ELS	218,000	11,900 - 118,500	4,276,000	4,505,900 - 4,612,500	8.7 - 11.2	1.5 - 0.2
58.6	ESP	S	SUB	228,000	11,900 - 118,500	3,804,000	4,043,900 - 4,150,500	1.1 - 3.8	0.2 - 0.1
58.6	ESP	I	EHS	114,000	11,900 - 118,500	4,113,000	4,238,900 - 4,345,500	5.0 - 7.7	0.9 - 0.1
58.6	ESP	I	ELS	206,000	11,900 - 118,500	4,276,000	4,493,900 - 4,600,500	8.4 - 11.0	1.5 - 0.2
58.6	ESP	I	SUB	211,000	11,900 - 118,500	3,804,000	4,026,900 - 4,133,500	0.7 - 3.4	0.1 - 0.1

Note: FF = Fabric Filter

MC = Multitube Cyclone

ESP = Electrostatic Precipitator

EHS = Eastern High Sulfur Coal

ELS = Eastern Low Sulfur Coal

SUB = Subbituminous Coal

Cost is also shown in terms of \$/kg particulate removed and is estimated as:

$$\text{\$/kg Removed} = \frac{\left[\begin{array}{c}\text{Total annual cost of AFBC without}\\ \text{SO}_2\text{ control but with}\\ \text{particulate control}\end{array}\right] - \left[\begin{array}{c}\text{Total annual cost}\\ \text{of uncontrolled}\\ \text{conventional boilers}\end{array}\right]}{\text{kg of particulate removed per year}}$$

The underlying accuracy (±40 percent) of the cost estimating procedure must be considered in evaluating the tabulated results. For instance, differences in cost between AFBC with particulate control and uncontrolled conventional boilers as a function of coal type may be as much a function of basic differences in boiler cost as particulate control device cost. Therefore, the impact of coal type on control device cost can only be determined from the first column of the table. ESP cost increases as coal sulfur decreases, but this trend would not be concluded from the last columns in the table.

Cost in terms of \$/kg particulate removed decreases as inlet loading increases from 215 to 2150 ng/J (0.5 to 5.0 lb/10^6 Btu). The range for ESP's may not be as wide as shown considering that, in reality, total ESP cost would increase with particle loading, but this analysis only accounts for the added cost of additional waste solids handling.

Fabric filters are cost-effective for stringent or intermediate control when low sulfur coals are burned. However, ESP's appear to have a cost advantage over fabric filters when high sulfur coal is burned. (There is still some question as to the performance of ESP's with FBC fly ash.) Considering equivalent boiler capacities, cost in terms of percentage increase and \$/kg particulate removed are not significantly different for either control device.

The percentage increase in cost over uncontrolled conventional boilers ranges as high as 20 percent for a boiler capacity of 22 MW_t. It decreases slightly with larger boiler sizes and is much lower for the 8.8 MW_t boiler simply because of the basic cost advantage of an AFBC boiler at this capacity. Cost in terms of $/kg particulate removed follows the same trend.

In several cases, negative values are shown in the last two columns of the table. This indicates that the cost of AFBC with the particulate control device noted and no SO_2 control is less expensive than an uncontrolled conventional boiler of the same capacity firing the same coal.

Based on final device cost alone, multitube cyclones are the cost-effective choice for moderate particulate control. However, the data do not show an overwhelming advantage when the cost of AFBC and final particulate control are considered together because variation in basic boiler cost tends to dampen the cost impact of particulate control application.

4.4.3 Cost of Particulate Control for AFBC Boilers - Including Influence of SO_2 Control

The combined cost of AFBC systems with SO_2 control and particulate control is shown in Table 45. Again, the costs are based on vendor quotes presented in the ITAR on particulate control.[42] The table assumes that within the accuracy of this study, control device performance on AFBC boilers with SO_2 control will be similar to conventional boilers burning low sulfur coal. Therefore, costs are presented based on estimates for conventional boilers burning subbituminous coal, the worst case cost. Consequently, all ESP costs represent hot side application.

TABLE 45. COST OF FINAL PARTICULATE CONTROL FOR COAL-FIRED AFBC INDUSTRIAL BOILERS WITH SO_2 CONTROL

Boiler capacity (MW_t)	Control device	Particulate control level	Final particulate control device cost			Annual[†] incremental spent solids disposal cost[‡]	Incremental spent solids handling and storage costs[§]		Total cost of final particulate control			Total annual cost of FBC with particulate control and SO_2 control	Percent increase in cost over uncontrolled conventional boilers
			Capital[*]	Operating	Annual[†]		Capital[*]	Annualized capital	Capital[*]	Operating	Annual[†]		
8.8	FF	S, I, M, SIP	238,000	13,000	51,000	2,700 - 13,600[#]	2,300 - 10,900[#]	300 - 1,600[#]	248,000 - 286,000[#]	14,600 - 21,200[#]	54,000 - 66,200[#]	911,000 - 1,090,000	(-2.4) - 17.6
22	FF	S, I, M, SIP	436,000	18,000	86,000	6,700 - 34,000	5,700 - 23,100	800 - 3,400	460,000 - 552,000	22,000 - 38,400	93,500 - 123,400	1,980,000 - 2,464,000	12.3 - 34.9
44	FF	S, I, M, SIP	766,000	27,000	147,000	13,400 - 68,000	10,800 - 65,500	1,600 - 9,600	814,000 - 1,018,000	35,000 - 67,800	162,000 - 226,600	3,220,000 - 4,229,000	6.3 - 38.9
58.6	FF	S, I, M, SIP	943,000	33,000	181,000	17,900 - 90,700	13,900 - 87,400	2,000 - 12,800	1,006,000 - 1,279,000	43,700 - 87,400	200,900 - 284,500	4,080,000 - 5,442,000	2.0 - 36.8
8.8	MC	M, SIP	51,000	2,000	10,000	2,700 - 13,600	2,300 - 10,900	300 - 1,600	60,700 - 99,200	3,600 - 10,200	13,000 - 25,200	870,000 - 1,049,000	(-6.8) - 13.2
40	MC	M, SIP	185,000	14,000	48,000	12,200 - 61,800	9,900 - 57,800	1,400 - 8,400	218,000 - 412,000	23,300 - 53,100	61,600 - 118,200	2,907,000 - 3,820,000	3.9 - 35.3
44	MC	M, SIP	98,000	11,000	26,000	13,400 - 68,000	10,800 - 65,500	1,600 - 9,600	145,000 - 349,800	19,000 - 51,800	41,000 - 103,600	3,079,000 - 4,108,000	2.3 - 34.9
8.8	ESP	S	414,000	10,000	75,000	2,700 - 13,600	2,300 - 10,900	300 - 1,600	424,000 - 462,000	11,600 - 18,100	78,000 - 90,200	935,000 - 1,114,000	0.1 - 20.2
8.8	ESP & MC	I	336,000	10,000	63,000	2,700 - 13,600	2,300 - 10,900	300 - 1,600	346,000 - 384,000	11,600 - 18,100	66,000 - 78,200	923,000 - 1,102,000	(-1.2) - 18.9
8.8	ESP	SIP	105,000	5,000	22,000	2,700 - 13,600	2,300 - 10,900	300 - 1,600	115,000 - 153,000	6,600 - 13,100	25,000 - 37,200	882,000 - 1,061,000	(-5.6) - 14.6
22	ESP	S	825,000	17,000	147,000	6,700 - 34,000	5,700 - 23,100	800 - 3,400	849,000 - 941,000	21,000 - 37,400	134,500 - 184,400	2,041,000 - 2,525,000	15.7 - 38.3
22	ESP	SIP	260,000	8,000	49,000	6,700 - 34,000	5,700 - 23,100	800 - 3,400	284,000 - 376,000	12,000 - 28,400	56,500 - 86,400	1,943,000 - 2,427,000	10.2 - 32.9
44	ESP	S	1,143,000	28,000	208,000	13,400 - 68,000	10,800 - 65,500	1,600 - 9,600	1,191,000 - 1,395,000	36,000 - 68,800	223,000 - 285,600	3,281,000 - 4,290,000	8.3 - 40.9
44	ESP & MC	I	1,214,000	35,000	226,000	13,400 - 68,000	10,800 - 65,500	1,600 - 9,600	1,262,000 - 1,466,000	43,000 - 75,800	241,000 - 303,600	3,299,000 - 4,308,000	8.9 - 41.5
44	ESP	SIP	865,000	17,000	153,000	13,600 - 68,000	10,800 - 65,500	1,600 - 9,600	913,000 - 1,117,000	25,000 - 57,800	168,000 - 230,600	3,226,000 - 4,235,000	6.5 - 39.1
58.6	ESP	S	1,249,000	31,000	228,000	17,900 - 90,700	13,900 - 87,400	2,000 - 12,800	1,312,000 - 1,585,000	41,700 - 85,400	247,900 - 331,500	4,127,000 - 5,489,000	3.2 - 36.0
58.6	ESP	I	1,162,000	28,000	211,000	17,900 - 90,700	13,900 - 87,400	2,000 - 12,800	1,225,000 - 1,498,000	38,700 - 82,400	230,900 - 314,500	4,110,000 - 5,472,000	2.8 - 35.6
58.6	ESP	SIP	1,008,000	20,000	179,000	17,900 - 90,700	13,900 - 87,400	2,000 - 12,800	1,701,000 - 1,344,000	30,700 - 74,400	198,900 - 282,500	4,078,000 - 5,440,000	2.0 - 34.8

[*] All capital costs are turnkey costs.

[†] Annual cost includes operating cost and annualized capital cost.

[‡] Disposal cost (based on $40/ton) of fly ash/sorbent captured in final control device. Amount captured is based on total system spent solids/ash quantities minus that amount withdrawn from combustor, which is included in the cost of SO_2 control.

[§] The same costing procedures and unit capital costs discussed in the subsection for SO_2 control are used here.

[#] Range represents the extremes in cost; the low being Eastern low sulfur coal with moderate SO_2 control and the particulate control level noted; the high being Eastern high sulfur coal with stringent SO_2 control and the particulate control level noted.

Control device/control level combinations are shown which were considered as "best system" candidates in Section 3.0. ESP costs at an SIP level are also shown for comparison, but the multitube cyclone is considered to be appropriate for SIP control. Percentage cost increases of implementing more stringent control than SIP is not shown because ESP use is not recommended at this low control level. Cost in terms of $/kg particulate removed should be similar to the values noted earlier in Table 49 for low sulfur coal.

In general, the total cost of an AFBC system with SO_2 and particulate control can range as high as 40 percent greater than a conventional boiler without any emission control. Fabric filters may be more cost-effective than ESP's in all cases for stringent or intermediate particulate control, since ESP's have been considered as hot side installations when SO_2 control is practiced.

This cost advantage is illustrated in Figure 49 which shows the cost of add-on particulate control devices. Depending on performance capability, cold side ESP's could be cost-effective compared to fabric filters. However, ESP's will probably not be capable of operating as cold side installations when SO_2 control is practiced in AFBC. Figure 49 also illustrates that multitube cyclones are the device of choice for moderate particulate control.

If inlet particle loadings are minimal (215 ng/J) and low sulfur coal is burned, the analysis indicates that the 8.8 MW_t AFBC boiler can be used at equal or less cost than an uncontrolled conventional boiler. This continues the trend shown in the SO_2 control cost analysis and is probably a result of the fairly low basic AFBC boiler cost at this capacity. This possible advantage must be confirmed in actual practice.

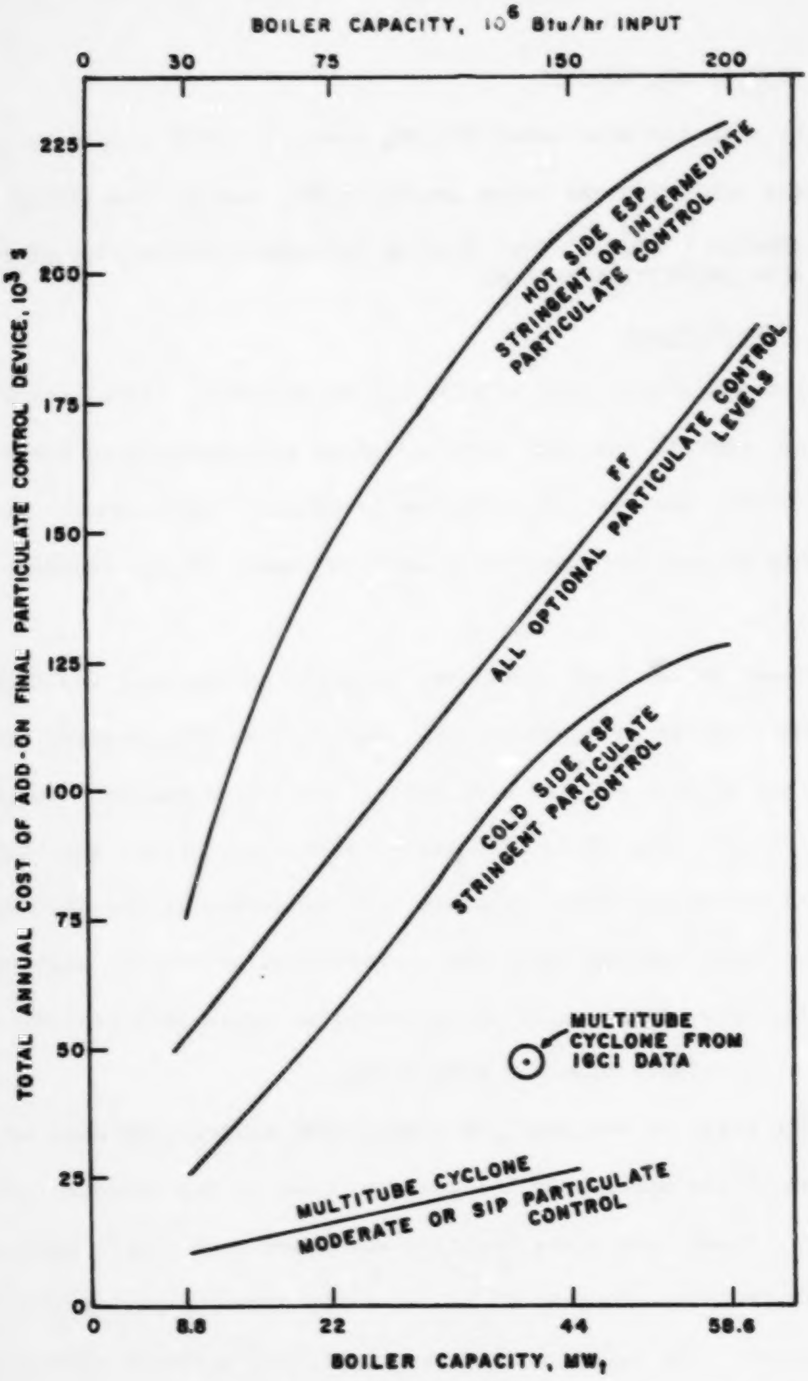


Figure 49. Cost of final particulate control for AFBC industrial boilers.

4.5 COST OF NO_x CONTROL

No cost has been added for NO_x control. AFBC should be capable of inherently achieving the three levels of NO_x control considered in this study.

4.6 SUMMARY - COST OF BEST SYSTEMS EMISSION CONTROL IN COAL FIRED AFBC INDUSTRIAL BOILERS

4.6.1 SO_2 Control

The annualized cost of AFBC boiler purchase, installation, and operation with SO_2 control has been computed along with comparable costs of uncontrolled conventional boilers (as estimated by PEDCo). This summary discusses the validity of the cost basis used and its impact on the accuracy of the final estimates.

Based on the cost quotations supplied by vendors, small AFBC boiler use (8.8 MW_t) can be of equal or less cost than an uncontrolled conventional stoker. AFBC cost becomes less as coal sulfur content decreases. In the larger capacities (22 to 58.6 MW_t), AFBC costs (with SO_2 control) are higher than uncontrolled conventional boiler costs. An exception is the 58.6 MW_t AFBC boiler burning subbituminous coal with a sorbent of average or high reactivity. In this instance, the cost of an uncontrolled pulverized coal-fired boiler is equal to or higher than the AFBC boiler.

The basis of the small (8.8 MW_t) AFBC boiler cost must be discussed because of its apparently low cost relative to the uncontrolled conventional boiler. First, the costs reported are based on a single basic boiler quote. The manufacturer (Company B) is currently offering package boilers in this size range. The boiler design is simple, but operates effectively based on demonstration plant operation over the last several months. Therefore, the costs presented are considered realistic.

One aspect of this AFBC boiler which is open to question with regard to "best" SO_2 control is the use of overbed screw feeding of coal and sorbent. To date, the SO_2 control capabilities of this technique in commercial operation are unknown, although tests indicate SO_2 can be effectively controlled. It is assumed for the purpose of this analysis that overbed feed can provide suitable SO_2 removal performance and that its cost is representative of "best system" cost. (See discussion of FluiDyne testing in Section 2.0 and 7.0).

In this report, we have attempted to indicate the full range of AFBC cost based on differences in sorbent reactivity, SO_2 control level, coal type, sorbent cost, and spent solids disposal costs. Because no large units have operated, the possible trade-offs between capital cost for optional feed systems and operating costs for reduced sorbent requirements cannot be quantified with total reliability, but rather, must be projected based upon small-scale experimental results and modeling efforts. In all probability, the added capital cost of in-bed materials feeding is within the worst case cost presented for AFBC with SO_2 control. Unless overbed screw feeding is proven inferior with respect to AFBC SO_2 control, there is no reason to modify the costs presented here.

The costs presented for the three larger boilers are also based on overbed coal feeders. The design in the larger boilers is somewhat different than that incorporated in the small system, but similar considerations apply with regard to SO_2 removal capabilities. The overbed feeding technique is under evaluation at Georgetown University.

The cost analysis indicates that AFBC with SO_2 control can cost up to 30 percent more than an uncontrolled conventional boiler. The maximum cost differential occurs at a stringent SO_2 control level during high sulfur coal combustion with a low reactivity sorbent. As coal sulfur content decreases, and SO_2 control level becomes more moderate, and as sorbent reactivity increases, the difference in cost between the technologies narrows significantly. AFBC was found to have equal or less cost at a capacity of 8.8 MW_t for either low sulfur coal and for Eastern high sulfur coal at an SIP control level. Also, the cost of AFBC was comparable or lower for the 58.6 MW_t AFBC burning subbituminous coal. These similarities must be verified after more thorough marketing and system use. Table 46 summarizes the cost of AFBC and uncontrolled conventional systems estimated in this study.

TABLE 46. COST SUMMARY - AFBC AND UNCONTROLLED CONVENTIONAL BOILERS: COST = $\$/10^6$ Btu OUTPUT

Coal type	Boiler type	Boiler capacity, MW_t			
		8.8	22	44	58.6
Eastern High Sulfur	AFBC	6.93 - 8.04	6.19 - 7.28	5.10 - 6.19	4.90 - 5.97
	Conventional	7.39	5.76	4.77	4.56
Eastern Low Sulfur	AFBC	6.79 - 6.93	6.13 - 6.27	5.06 - 5.19	4.86 - 4.99
	Conventional	7.12	5.62	4.70	4.55
Subbituminous	AFBC	6.66 - 6.79	5.80 - 5.93	4.68 - 4.80	4.44 - 4.56
	Conventional	7.41	5.54	4.73	4.57

An important conclusion of this study is the apparently small cost difference between removing 75 or 90 percent SO_2 using AFBC. The greatest difference occurs for high sulfur coal combustion (~$\$0.30/10^6$ Btu for average sorbent reactivity) but the difference becomes insignificant for low sulfur coals. Sorbent reactivity can have a larger cost effect than control level depending on the extremes in reactivity considered.

Implementation of "best system" conditions for SO_2 control can reduce the cost of FBC compared to "commercially offered" design/operating conditions. This is mainly due to reduced operating costs. Capital costs may be higher or lower depending on the alterations necessary and the specific design of interest.

4.6.2 Comparison with FGD

Considering the accuracy of both conventional and AFBC boiler costs presented in this report, it is difficult to draw clear cut conclusions concerning the cost-effectiveness of SO_2 control employing AFBC technology. Comparison with preliminary flue gas desulfurization (FGD) costs prepared by Radian[43] for coal-fired industrial boilers can lend some perspective to the results of the AFBC cost analysis. Table 47 lists the costs of FGD and AFBC in terms of percentage increase over the cost of uncontrolled conventional boilers. For the FGD case, the reported ranges cover low and high sulfur coals and optional levels of SO_2 control. The AFBC ranges include, in addition, extremes in sorbent reactivity. The data indicate that AFBC has a cost advantage at a boiler capacity of 8.8 MW_t, but that the maximum cost of both technologies becomes comparable as boiler capacity increases from 22 up to 58.6 MW_t. On this basis, it is concluded that AFBC is a cost-effective SO_2 control technology and that it should be considered in any instance where SO_2 control is required for coal-fired industrial boilers.*

4.6.3 Particulate Control

The results of the particulate control cost analysis (estimated accuracy = ±40 percent) indicate that fabric filters or ESP's may be selected for stringent or intermediate control depending on coal type and implementation of SO_2 control. Without SO_2 control, the estimated ESP costs are based on cold side installation when high sulfur coal is burned. Under this condition ESP's are less expensive than fabric filters. For any other condition; i.e., low sulfur coal or inclusion of SO_2 control, fabric filters appear to be cost-effective.

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*Lack of full scale operating data is still the major bottleneck in the technology's development.

TABLE 47. RELATIVE COMPARISON OF THE COST OF AFBC VERSUS CONVENTIONAL BOILERS WITH FGD

FGD process	Boiler capacity MW_t	% Increase in cost over uncontrolled conventional boilers[*]	
		FGD[†]	AFBC
Limestone	8.8	35 - 46	<10
	22	25 - 37	7 - 29
	58.6	17 - 26	<28
Sodium	8.8	32 - 44	<10
	22	23 - 38	7 - 29
	58.6	16 - 32	<28
Double Alkali	8.8	35 - 46	<10
	22	24 - 37	7 - 29
	58.6	17 - 27	<28
Wellman-Lord	8.8	36 - 51	<10
	22	25 - 41	7 - 29
	58.6	18 - 29	<28

[*] Range includes low and high sulfur coals and optional SO_2 control levels. For AFBC, extremes in sorbent reactivity are also included.

[†] Based on Radian TAR on FGD; see Reference No. 45.

Under the more realistic condition where SO_2 control is assumed, fabric filters seem to be the control device of choice, considering potential problems with particle resistivity in ESP's and the loss of normally condensable trace elements during hot side control. However, potential problems with fabric filter use, such as blinding or bag fires, must be assessed in commercial operation before one technique can be recommended over the other with total confidence.

For moderate particulate control, multitube cyclones are the cost-effective choice based on this analysis. It is important to reiterate that the accuracy of the estimating technique is limited and that results must be verified in actual applications.

4.6.4 NO_x Control

NO_x control to the three levels considered in this report of 215, 258, 301 ng/J (0.5, 0.6, 0.7 lb/10^6 Btu) is assumed to be inherently achievable in AFBC. Therefore, no costs have to be added for NO_x control.

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5.0 ENERGY IMPACT - FLUIDIZED-BED COMBUSTION VERSUS CONVENTIONAL BOILERS

5.1 INTRODUCTION

The objective of this section is to quantify the energy impact of pollution control in an atmospheric fluidized-bed combustor when compared with the power requirements of standard uncontrolled conventional boilers.

The pollutants controlled are SO_2, NO_x, and particulates. The inherent chemistry of fluidized-bed combustion results in sufficiently low NOx emissions that no energy penalty for NO_x control is expected. Because particulate emissions from the two technologies should be similar, (see Section 2.0) energy requirements for AFBC particulate control are estimated based on conventional firing control technology.[1] Most of this discussion addresses the energy impact of SO_2 control in fluidized-bed combustion.

A qualitative comparison of uncontrolled conventional firing, AFBC, and conventional firing with wet scrubbing is presented in Table 48. Several items of energy use common to all systems and of similar impact are noted. Important energy impacts associated with flue gas desulfurization which are not a factor in FBC are liquid pumping through the scrubber loop, absorption tower pressure drop, and flue gas reheat.

Performance of a mass and energy balance around both an AFBC and a conventionally-fired design permits quantification of the energy requirements for

TABLE 48. QUALITATIVE COMPARISON OF ENERGY IMPACT ASSOCIATED WITH AFBC AND CONVENTIONAL COAL-FIRED INDUSTRIAL BOILERS

Subsystem	Components of energy use	AFBC with control	Conventional without control	Conventional with FGD	Other comments
Coal handling	Conveying and feeding	*	*	*	AFBC has advantage of lower crushing power needs, in comparison to pulverized coal-firing, due to ability to feed coarser coal.
	Crushing and sizing	Lower energy requirements than conventional systems	*	*	
	Drying	*	*	*	
Limestone handling	Conveying Screening	Small power requirements; for worst case, less than 0.1% of thermal input to boiler.	NA	Lower power consumption than AFBC for all components because of lower sorbent loadings due to lower required Ca/S ratio.	
	Calcination Sulfation	Depending on sorbent reactivity, energy added by sulfation can approach or outweigh loss due to calcination.			
Spent solids/ash handling	Conveying Cooling		Lowest		
Forced draft fan	Air heater	*	*	*	The largest auxiliary power requirement in AFBC is forced draft fan operation. Assuming, 20% excess air and Δp noted, FD fan power is equivalent to 1.0% of total thermal input to boiler.
	Ducts and steam coil heater	*	*	*	
	Plenum	*	*	*	
	Burners	NA	*	*	
	Distributor plate	High auxiliary power requirements, average total pressure loss of 140 cm (55 in.) w.g.	NA	NA	
	Fluid bed		NA	NA	
Induced draft fan	Freeboard	Equivalent to conventional furnace Δp	NA	NA	ID fan power requirements are generally higher for AFBC operation due to inclusion of primary cyclone for char and sorbent recirculation and reduction of particulate loading to final control device.
	Furnace	NA	*	*	
	Boiler and superheater	*	*	*	
	Primary cyclone	Maximum Δp = 15 cm (6 in.) w.g.	NA	NA	
	Economizer	*	*	*	
	Air heater	*	*	*	
	Flues	*	*	*	
Boiler feedwater system	Pumps and other equipment Chemical feed Heating Blowdown	*	*	*	All systems have similar energy requirements.
Final particulate control	Control device Operating power	*	*	*	Considering ESP use, it is probable that hot side application will be required in more cases with AFBC than conventional boilers. This gives a slight energy advantage to conventional boilers.
Wet scrubber plant	Pumping Absorber tower Flue gas reheat	NA	NA	High auxiliary power requirements in the range of 2.0% of total thermal input to boiler. Flue gas reheat is a large portion of the total auxiliary power requirement.	
Sensible heat loss	Spent solids/ash	High energy loss compared to conventional with and without scubbing due to large quantities of limestone.	lowest because only component is coal ash	Intermediate due to scrubber sludge loss along with bottom ash loss.	
	Flue gas	Lowest because of low excess air relative to stokers.	intermediate because of excess air losses	Highest losses due to excess air and reheat requirements after wet scrubbing.	
Unburned carbon loss	Elutriation Bottom ash Fly ash	*	*	*	Pulverized coal firing has demonstrated 99+% carbon utilization.
Radiative and other unaccounted losses	Boiler surface piping	*	*	*	This loss becomes successively less of a percentage of the total input as boiler size increases due to lower surface-to-volume ratios.

[*] Indicates similar energy requirements or losses for different systems.

NA = Not applicable.

both designs.* Each unit operation within the system was evaluated and the loss component assessed. The detailed tables derived are presented in Appendix C. The losses associated with each operation were grouped in terms of auxiliary or inherent losses. Auxiliary losses are those deriving from electric power requirements for process operations. Inherent losses are the sensible heat losses, heat of reaction losses, and phase change losses.

Important energy losses in AFBC boilers are: air pressure drop across the combustion air distribution plate, fluid bed, and primary cyclone; limestone calcination; flue gas sensible heat loss; unburned carbon loss; solids conveying; and spent solids sensible heat loss. A schematic diagram of a standard AFBC industrial boiler system is shown in Figure 50, and illustrates the auxiliary equipment necessary for SO_2 and particulate control.

In the following subsections, the energy impacts of AFBC operation are itemized. The total energy impact of SO_2 reduction via AFBC is derived as a function of SO_2 control level, standard boiler capacity, sorbent reactivity, and coal characteristics. Ultimately, the increase in energy use over the uncontrolled standard conventional boilers is presented along with a parametric sensitivity analysis.

The results of these energy analyses indicate that energy penalty for SO_2 control is mainly a function of boiler size. Large boilers firing

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*For this study, fluidized-bed combustion and conventional coal-fired boilers having no SO_2 control are compared. For perspective, a comparison of a fluidized-bed combustion and a conventional boiler system incorporating flue gas desulfurization is made later in this section.

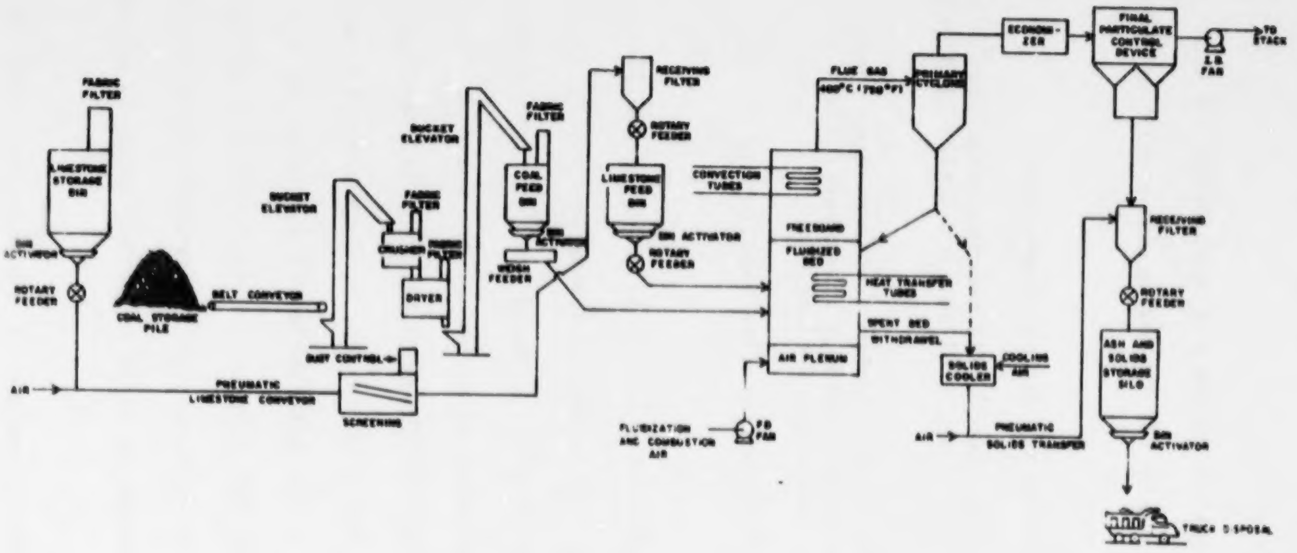
FABRIC FILTER
LIMESTONE STORAGE BIN
BIN ACTIVATOR
ROTARY FEEDER
AIR
COAL STORAGE PILE
BELT CONVEYOR
PNEUMATIC LIMESTONE CONVEYOR
BUCKET ELEVATOR
FABRIC FILTER
CRUSHER
FABRIC FILTER
DRYER
DUST CONTROL
SCREENING
BUCKET ELEVATOR
FABRIC FILTER
COAL FEED BIN
BIN ACTIVATOR
WEIGH FEEDER
RECEIVING FILTER
ROTARY FEEDER
LIMESTONE FEED BIN
BIN ACTIVATOR
ROTARY FEEDER
CONVECTION TUBES
FREEBOARD
FLUIDIZED BED
HEAT TRANSFER TUBES
SPENT BED WITHDRAWAL
AIR PLENUM
FLUIDIZATION AND COMBUSTION AIR
FD FAN
FLUE GAS 400°C (750°F)
PRIMARY CYCLONE
SOLIDS COOLER
COOLING AIR
AIR
PNEUMATIC SOLIDS TRANSFER
ECONOMIZER
FINAL PARTICULATE CONTROL DEVICE
I.D. FAN
TO STACK
RECEIVING FILTER
ROTARY FEEDER
ASH AND SOLIDS STORAGE SILO
BIN ACTIVATOR
TRUCK DISPOSAL

Figure 50. Schematic of AFBC industrial boiler including auxiliary equipment (assumes carbon burnup cell will not be necessary).

pulverized coal are more efficient than the AFBC furnaces; while the stoker-fired furnaces are less efficient. Other variables which affect efficiency are coal type and sorbent reactivity.

5.2 AUXILIARY EQUIPMENT ENERGY DEMAND FOR SO_2 CONTROL IN AFBC

Auxiliary power is used by the following components in AFBC boilers with SO_2 control:

1. Coal handling (crushing, drying, conveying).
2. Boiler feedwater treatment, all pumping associated with feedwater, condensate and cooling circulation, and other miscellaneous systems.
3. Forced draft fan, induced draft fan (excluding power needs for pressure drop through final particulate control device), pneumatic feeding, etc.
4. Limestone handling, spent solids handling.

5.2.1 Coal Handling

Power requirements associated with coal handling include crushing, sizing, drying, and conveying. In this study, crushing and sizing are assumed to be performed in the same process module, while conveying and drying require additional modules. Estimated power requirements are based on relationships discussed in Perry's Handbook of Chemical Engineering, Sections 7, 8, and 20.[2]

In the crushing and sizing operation, coal is assumed supplied run-of-mine (-6 in.). Required feed to the AFBC boiler is specified as 2.5 cm (-1 in.) and under. The pulverized coal furnace requires -74 µm (-200 mesh) and the stokers require -2.5 cm (-1 in.). Power requirement estimates are based on the assumption that horsepower is directly proportional to reduction ratio and capacity.[3]

Coal drying to a moisture content of 5 percent is required for any system using pneumatic coal feeding. The stokers and the AFBC designs do not require drying. The pulverized coal-fired boiler is the only unit where coal drying is required. To meet a 5 percent moisture limit, 3.79 percent moisture must be removed from the Eastern high sulfur coal and 15.8 percent moisture must be removed from the subbituminous coal. No drying is required for the Eastern low sulfur coal since the as-received moisture content is below 5 percent.

A fluidized-bed dryer was chosen for this study.* One of the major advantages of this type of dryer in coal drying is the close control of conditions so that a predetermined amount of free moisture may be left with the solids to prevent dusting during subsequent operations. Fuel requirements are from 1500 to 1900 Btu/lb of water removed and total power for blowers, feeders, and related equipment is about 0.037 kW-hr/lb of water removed.[4] For this study an average requirement of 1700 Btu/lb of water removed was assumed. Heat for drying is supplied by the boiler.

A point worth noting is that moisture not removed during the drying operation results in a flue gas latent heat loss of 1040 Btu/lb of water plus the sensible heat loss. Thus, while the drying requirement results in significant increases in coal handling energy penalties, this loss is balanced by somewhat reduced flue gas losses.

Energy requirements for conveying, include power to move coal from storage, between process modules, and to the primary fuel hopper. Conveying is done using belt, bucket, and flight conveyors and pneumatic equipment. Conveying power requirements are based on correlations presented in Perry's coupled with the total tonnage of material involved. Conveying power requirements include a 50 percent contingency factor to cover intermittant loads. This adds 2 kW to the small boiler energy loss and 10 kW to the largest boiler loss.

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*Although a ball mill would be used for crushing and drying in the pulverized case, the assumption of use of a fluidized bed dryer does not affect the accuracy of the estimating procedure used here. The important factor in the analysis is that some type of component is used do remove the level of moisture noted.

Table 49 summarizes auxiliary power required for coal handling in AFBC as a function of boiler capacity, and coal. A comparison is provided with the auxiliary requirements of the most likely competitive conventional system in each of the respective size ranges.

5.2.2 Boiler Feedwater Treatment and Auxiliary Pumping Requirements

Power required for boiler feedwater treatment and all necessary pumping is considered to be a function of boiler capacity only. Energy requirements listed in Table 50 are based on forced circulation boiler pumping requirements plus a 15 percent contingency to cover small and intermittent loads. These power requirements are extrapolated from estimates for a forced circulation boiler by Babcock and Wilcox Company.[5] A forced circulation design was estimated because many designs for FBC require forced circulation. If natural convection proves feasible, pumping energy requirements can be reduced.

5.2.3 Forced Draft and Induced Draft Fan Power

Forced draft (FD) and induced draft (ID) power represents the largest electrical consumption in AFBC operation. The FD fan must be of sufficient capacity to move air through the air heater, ducting, plenum, distributor plate, and fluid bed. The ID fan must transport flue gas from the freeboard, through the primary cyclone, economizer, air heater, and flue. (Power required for flue gas movement through the final particulate control device is discussed later.) Table 51 shows total AFBC fan power requirements for combustion and SO_2 removal as a function of boiler capacity. Fan power in conventional systems is also shown for comparison. For a detailed breakdown of the components considered, see Appendix C, Table C-5.

TABLE 49. AUXILIARY ENERGY[*] REQUIRED FOR COAL HANDLING

Boiler capacity MW_t (10^6 Btu/hr)	Burner type	Auxiliary energy - KW		
		Eastern high sulfur coal	Eastern low sulfur coal	Subbituminous coal
8.8 (30)	Stoker[†]	6	5	7
	AFBC	6	5	7
22 (75)	Stoker[†]	12	11	14
	AFBC	12	11	14
44 (150)	Stoker[†]	22	19	27
	AFBC	22	19	27
58.6 (200)	Pulverized[†]	373	25	1796
	AFBC	29	25	35

[*] GCA estimates.

[†] Uncontrolled.

TABLE 50. AUXILIARY POWER* REQUIRED FOR BOILER FEEDWATER CIRCULATION, TREATMENT AND ALL ASSOCIATED PUMPING IN CONVENTIONAL AND AFBC

Boiler capacity MW_t (10^6 Btu/hr)	Auxiliary power KW (HP)
8.8 (30)	18 (25)
22 (75)	47 (63)
44 (150)	94 (125)
58.6 (200)	125 (167)

[*] GCA estimates.

TABLE 51. AUXILIARY POWER* FOR FORCED DRAFT, INDUCED DRAFT, AND ANCILLARY AIR

Boiler capacity MW_t (10^6 Btu/hr)	Burner type	Auxiliary power (kW)	Flue gas rates (acfm)
8.8 (30)	Stoker†	42	12,500
	AFBC	115	10,000
22 (75)	Stoker†	91	31,400
	AFBC	287	25,120
44 (150)	Stoker†	172	62,800
	AFBC	574	50,240
58.6 (200)	Pulverized†	277	73,200
	AFBC	766	67,570

*GCA estimates.

†Uncontrolled.

Pressure losses through the economizer and other common equipment components were estimated by reference to Steam/Its Generation and Use, by Babcock and Wilcox.[6] Pressure loss through the FBC distribution plate and fluid bed was estimated by reference to experimental data reported by Pope, Evans, and Robbins.[7] For plate designs tested, the average pressure loss equaled twice the velocity head. Assuming a range of superficial gas velocities in industrial AFBC boilers between 1.8 to 2.4 m/sec (6 to 8 ft/sec), a representative loss through the distribution plate is 38.1 cm (15 in.). Pressure loss (w.g.) in the bed during PER testing was found to be approximately equal to the expanded bed height.[8] In this analysis, a bed depth of 122 cm (48 in.) is assumed for estimating FBC FD fan power, in conformance with the bed depth recommended for "best system" design.

The selection of this bed height represents a compromise between two factors. First, increased bed depth results in increased pressure drop, which puts more load on the forced draft fan. Conversely, decreasing the bed height will result in lower sorbent and gas residence times with concomitant increases in either sulfur emissions or sorbent requirements.

This interrelation between bed depth and sorbent requirement may severely limit the application of bed height variation as a method of load following (see Section 2.0). If bed height variation is attempted as a load following technique, bed depths lower than 30 in. are possible. The lower value will depend upon tube surface area which must be exposed to achieve the desired boiler turndown. An important point to note is that this shallow bed will have severely impaired sulfur capture capability and could not be maintained without penalties in SO_2 emissions or sorbent requirements.[9] It seems likely that bed slumping, variation in superficial velocity, and bed temperature control will be more acceptable methods of load following.

While no one bed height will serve in all designs, an estimate of 122 cm (48 in.) should be representative for conventional AFBC designs. In cases where sorbent is expensive, or of low reactivity, the additional fan loss associated with increased bed depth (to obtain higher sorbent sulfation and higher combustion efficiency) may be acceptable.

Flue gas rates required for calculating fan power requirements for conventional boilers are average figures (i.e., subbituminous coal-firing) taken from PEDCo reference data.[10] Flue gas rates for AFBC were proportioned from the conventional boiler estimates, assuming 20 percent excess air in all four standard AFBC boilers. Combustion air rates for both systems were estimated assuming a temperature of 22°C (80°F) for forced draft fan design. Fan power was estimated using standard design practice and a fan efficiency of 65 percent.[11]

Total fan power requirements for AFBC with SO_2 control are about three times that necessary for conventional boiler operation. AFBC fan power ranges from 115 to 766 kW for boilers ranging in capacity from 8.8 to 58.6 MW_t (30 to 200×10^6 Btu/hr). These figures represent the calculated power requirements plus a 10 percent contingency to cover ancillary air requirements.

5.2.4 Limestone and Spent Solids Handling

Limestone and spent solids handling auxiliary power requirements were estimated from the materials quantities coupled with the estimated unit power requirements (in kW/100 kg of solids) presented in Table 52. Power requirements for limestone and spent solids handling in Table 52 were determined by reference to a system (approximate coal-fired capacity equals 34 MW_t) under construction by Foster-Wheeler.[12]

TABLE 52. POWER USED FOR MATERIALS HANDLING IN AFBC COAL-FIRED BOILERS

Equipment item	Power use - KW (HP)	
	Limestone handling Feed Rate = 1,634 kg/hr (3,600 lb/hr)	Spent solids handling Withdrawal Rate = 1,362 kg/hr (3,000 lb/hr)
1. Fresh limestone air blower	7.5 (10)	
2. Rotary feeders	1.1 (1.5)	0.4 (0.5)
3. Limestone feed air blower	2.2 (3.0)	
4. Limestone screening (with dust control)*	1.9 (2.5)	
5. Solids cooler		3.0 (4)
6. Solids cooler fan		7.5 (10)
7. Spent solids air blower		18.6 (25)
8. Bin activators	3.0 (4.0)	1.5 (2.0)
9. Total	15.7 (21.0)	31.0 (41.5)
Unit power requirements kw/100 kg/hr (HP/100 lb/hr)	0.96 (0.58)	2.28 (1.38)

*Based on correspondence with C.E. Tyler Elevator Products. (See discussion in text.)

Materials quantities are a function of boiler size, coal type, SO_2 control level, and sorbent reactivity. The Ca/S molar feed ratios used in Table C-6, Appendix C, are based on the test data presented in Section 3.0. A range of Ca/S ratios is considered for each coal and each control level, assuming a range of sorbent reactivities. Limestone is assumed to be 90 percent $CaCO_3$, with 95 percent calcination to CaO. Spent bed material quantities include limestone inerts, uncalcined limestone, unreacted CaO, $CaSO_4$ generated and coal bottom ash. (The exact method of calculating spent solids quantities is shown in Section 6.0, Table 20).

The screening power requirements noted in Table 52 are based on correspondence with C.E. Tyler Elevator Products of Mentor, Ohio.[13] Although limestone conveying and spent solids handling needs can be represented readily, limestone crushing and screening requirements are difficult to characterize on a general basis for two reasons. First, the particle size distribution of limestone received from the quarry is variable from quarry to quarry. Second, because the physical characteristics of different limestones are variable, the ultimate limestone particle size distribution in the bed will be affected by attrition and elutriation. In some instances, an appropriate particle size distribution (average size of ~500 μm) may be attained in the bed with no intermediate processing required at the quarry or industrial site. In the extreme case, crushing and screening may be necessary. In any event, limestone crushing and screening requirements will be determined on a case-by-case basis.

To estimate auxiliary energy requirements for limestone processing at the FBC site, using the input from C.E. Tyler Elevator Products, double screening is assumed at the FBC site, but all crushing is performed at the quarry. Power is utilized in screening for mechanical vibration and for fan operation to convey entrained dust through a hood and cyclone or fabric collector. For processing of 1,634 kg/hr (3,600 lb/hr) limestone, total power for screening and dust control is estimated at 1.9 kW (2.5 hp).[14]

Unit power requirements for materials handling, as shown in Table 52, were applied to the full range of limestone and spent solids rates. Table 53 indicates the range of total materials handling power use as a function of boiler capacity and firing method. Table 53 is a summary of Table C-6 which details the complete range. Materials handling power requirements are maximum at the highest sorbent feed rate; i.e., burning high sulfur coal at a stringent SO_2 control level using a sorbent of low reactivity. For a particular coal, variation in materials handling power is most dependent on sorbent reactivity.

5.2.5 Total Auxiliary Power Requirements

The various electrical loads identified in the previous subsections are summed and presented in Table 54 as a function of coal grade, control level and sorbent reactivity for the firing methods and boiler sizes considered.

Because this represents the purchased electric power requirements in an industrial boiler, the heat supplied by the boiler for coal drying was subtracted from the total in Table 54 and added to the inherent losses in Section 5.3. Auxiliary power requirements for AFBC are higher than the auxiliary power requirements for uncontrolled conventionally-fired boilers. (Auxiliary power estimates for conventional units with SO_2 control would be somewhat higher than the uncontrolled units.) The chief component of this loss differential is the fan power requirements which represent roughly 60 percent of the total auxiliary power purchased in a conventional system and 70 percent for AFBC.

TABLE 53. AUXILIARY POWER* REQUIRED FOR CONVENTIONAL AND AFBC SOLIDS HANDLING

Boiler type	Boiler capacity MW_t (10^6 Btu/hr)	Auxiliary power - KW
		All coal types
Stoker	8.8 (30)	2 - 3[†]
AFBC		3 - 19
Stoker	22 (75)	4 - 7
AFBC		8 - 48
Stoker	44 (150)	9 - 14
AFBC		16 - 96
Pulverized coal	58.6 (200)	12 - 19
AFBC		22 - 128

[*] GCA estimates.

[†] The range presented represents variability to expect when going from Moderate control with a high reactivity sorbent to Stringent control with a low reactivity sorbent.

TABLE 54. TOTAL AUXILIARY POWER REQUIREMENTS FOR AFBC AND UNCONTROLLED CONVENTIONAL BOILERS — kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	70.	155.	157.	386.	302.	771.	399.	1027.
		LOW	4.2	70.	158.	157.	394.	302.	786.	399.	1047.
		HIGH	2.3	70.	152.	157.	378.	302.	754.	399.	1004.
	I 85%	AVERAGE	2.9	70.	154.	157.	382.	302.	765.	399.	1017.
		LOW	3.8	70.	157.	157.	390.	302.	778.	399.	1037.
		HIGH	2.1	70.	151.	157.	376.	302.	750.	399.	999.
	M 78.7%	AVERAGE	2.5	70.	152.	157.	379.	302.	756.	399.	1007.
		LOW	3.4	70.	155.	157.	386.	302.	771.	399.	1027.
		HIGH	1.8	70.	150.	157.	373.	302.	744.	399.	991.
	SIP 56%	AVERAGE	1.0	70.	147.	157.	365.	302.	728.	399.	970.
		LOW	1.2	70.	147.	157.	366.	302.	731.	399.	974.
		HIGH	0.8	70.	146.	157.	363.	302.	724.	399.	965.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	68.	143.	152.	354.	293.	707.	387.	943.
		LOW	3.7	68.	143.	152.	356.	293.	711.	387.	947.
		HIGH	2.0	68.	142.	152.	353.	293.	705.	387.	939.
	M 75%	AVERAGE	2.2	68.	142.	152.	353.	293.	705.	387.	940.
		LOW	3.2	68.	143.	152.	355.	293.	709.	387.	944.
		HIGH	1.6	68.	142.	152.	352.	293.	703.	387.	937.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	69.	144.	157.	358.	302.	715.	399.	953.
		LOW	3.6	69.	145.	157.	360.	302.	718.	399.	957.
		HIGH	2.0	69.	144.	157.	357.	302.	713.	399.	950.
	M 75%	AVERAGE	2.2	69.	144.	157.	357.	302.	713.	399.	950.
		LOW	3.2	69.	144.	157.	359.	302.	717.	399.	955.
		HIGH	1.6	69.	143.	157.	356.	302.	711.	399.	948.

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5.3 INHERENT ENERGY LOSSES IN THE FBC SYSTEM

Energy losses (other than auxiliary power) associated with AFBC coal combustion are the heat losses in flue gas and spent solids, limestone calcination, unburned carbon, and radiative and convective losses. (The total inherent energy loss also includes the coal drying losses estimated in the previous subsection.) Each loss is quantified and the effects of design and operating variations are discussed.

5.3.1 Flue Gas Heat Loss

Flue gas heat loss represents the single largest loss associated with coal-fired steam production. The components of this loss are latent heat, sensible heat, and humidity. The magnitude of each component is a function of coal composition and moisture content, excess air, and temperature differential between ambient air and flue gas. The temperatures assumed in the analysis are: ambient - 27°C (80°F); conventionally-fired high sulfur flue gas - 200°C (400°F); conventionally-fired low sulfur and subbituminous flue gas - 175°C (350°F). Flue gas temperatures are assumed at 175°C (350°F) in all AFBC cases. Excess air rates of 50 percent for stoker-fired boilers, 30 percent for pulverized coal furnaces, and 20 percent for AFBC were used in this study. The conventional boiler excess air rates are taken from the PEDCo study.[15] The AFBC air rate is the mid-range commonly reported by vendors. Reduction of the excess air to 10 percent may be possible through improved design and two-stage combustion. (Two-stage operation is being investigated in Sweden by O. Mustad and Son.)

Coal composition and moisture content affect the sensible and the latent heat content of the flue gas. Coal analyses and moisture content are taken from the PEDCo study of conventionally-fired boilers[16] (see Table C-1), in

Appendix C). In the cases where coal drying is required, the flue gas sensible heat loss is reduced by the amount of moisture removed during drying. The results of the flue gas heat loss calculations are shown in Table 55.

5.3.2 Solids Heat Loss

The heat loss accompanying spent solids withdrawal is calculated using a standard heat balance of the form:

$$Q = W_{out} \cdot C_{p_{out}} \cdot (T_{out} - T_{in})$$

where the heat capacity (C_p) of spent bed material plus ash is 947 J/kg-°K.[17] The weight of the material out is represented by W, the temperature by T, and the heat loss by Q. A value of 947 J/kg-°K is also assumed for the ash in conventional boilers. The AFBC bed solids temperature differential is 1480°F and the conventional bottom ash temperature differential is 1700°F.

For AFBC, 90 percent of the input ash is retained as bed residue. The 8.8 MW_t, and 22 MW_t stokers retain 75 percent of the ash as bottoms, the 44 MW_t stoker retains 35 percent as bottoms, and the pulverized coal-fired 58.6 MW_t unit retains 20 percent. Even though some solids exit the system as bottom residue and other material exits with the flue gas, both stream solids losses (bottoms and elutriated) are reported in Table 56 as solids heat losses. The differentiation between retained solids and elutriated solids is necessary because of the temperature differences between solids in the bed and solids in the flue gas. Systems with higher entrainment rates have lower solids heat losses because of cooling and subsequent heat recovery from the solids and flue gases in the economizer.

In addition to the sensible heat loss in the FBC, both the endothermic limestone calcination reaction and the exothermic sulfation reaction must be accounted for. Calcination requires 3,178 kJ/kg per kg CaO produced and

TABLE 55. FLUE GAS HEAT LOSSES*

Boiler capacity MW_t (10^6 Btu/hr)	Burner type	Heat losses - KW Eastern high sulfur	 Eastern low sulfur coal	 Subbituminous coal
8.8 (30)	Stoker	1277	1065	1270
	AFBC	955	883	1074
22 (75)	Stoker	3192	2664	3176
	AFBC	2388	2207	2685
44 (150)	Stoker	6384	5327	6351
	AFBC	4777	4415	5370
58.6 (200)	Pulverized	7381	6317	6506
	AFBC	6369	5886	7160

*See Appendix C, Table C-1 for coal analyses on which heat loss calculations are based.

TABLE 56. ENERGY IMPACT OF SOLIDS HEAT LOSS (INCLUDES CALCINATION AND SULFATION REACTIONS FOR FBC)

Boiler type	Boiler capacity MW_t (10^6 Btu/hr)	Energy impact - kW* All coal types
Stoker	8.8 (30)	13 - 24
AFBC		1 - 213
Stoker	22 (75)	33 - 61
AFBC		3 - 533
Stoker	44 (150)	39 - 72
AFBC		6 - 1066
Pulverized coal	58.6 (200)	37 - 72
AFBC		8 - 1421

*Assumes no heat recovery from the withdrawn spent bed material.

sulfation of CaO releases 8,668 kJ/kg per kg CaO consumed.[18] This consideration provides further impetus for using highly reactive sorbents and low Ca/S ratios. In cases where the sorbent stone is highly sulfated, a net heat release for the two reactions can be achieved.

The solids heat balance is summarized in Table 56. (The complete table presenting all values is in Appendix C, Table C-10.) This table presents the range of values calculated when one considers moderate control with high reactivity sorbent through stringent control with low reactivity sorbent. When sensible heat, calcination, and sulfation are accounted for, energy losses range from 1 to 213 kW for the small boiler (8.8 MW_t) and 8 to 1421 kW for the larger boiler (58.6 MW_t).

Variables which will affect the total solids loss are: the quantity of ash and limestone input, the retention/elutriation split, flue gas and spent solids exit temperature, and the degree of calcination and sulfation achieved. The quantity of limestone required is a function of coal sulfur, SO_2 control level, and limestone reactivity. Selection of a reactive limestone and precise control of the Ca/S molar feed ratio will both serve to minimize these losses.

5.3.3 Combustion Losses

A wide range of combustion efficiencies has been reported for AFBC units: 85 to 90 percent for units operating without recycle of solids from the primary cyclone and 95 to 97 percent for units operating with recycle.[19,20,21] Convention-firing combustion efficiencies range from 95 to 97 percent for spreader stokers with recycle. Pulverized coal units (the 58.6 MW_t conventional case) have demonstrated the capability of routinely achieving 99+ percent combustion efficiency.

For this study the upper end of the reported range, 97 percent, was assumed achievable for both spreader stoker and AFBC boilers. A combustion efficiency of 99 percent was assumed for the pulverized coal-fired unit. Table 57 presents the combustion loss estimates based on the efficiencies noted.

TABLE 57. COMBUSTION LOSS

Boiler capacity MW_t (10^6 Btu/hr)	Energy loss - kW	
	Conventional	AFBC
8.8 (30)	264	264
22 (75)	659	659
44 (150)	1,318	1,318
58.6 (200)	586	1,757

The combustion efficiencies assumed can be achieved through both good design practice and good operating procedures. Recent AFBC designs, for example, have higher freeboards than earlier systems. This higher freeboard improves combustion efficiency, probably by reducing char elutriation. Increasing gas residence time with deeper beds and lower superficial velocities as recommended for improved sulfur retention also serves to improve combustion efficiency.

Operator-controlled variables which affect combustion efficiency are the ratio of char recycle to char rejection, coal sizing, and the superficial velocity. Recycle of a large percentage of the elutriated material will increase carbon burnout while increasing the load on the particulate control device. Rejection of coal fines will reduce the char elutriation problem while increasing coal costs. Low superficial velocities will reduce solids carryover while requiring a larger boiler size for a given steam output. Thus, each option for improved carbon burnout is accompanied by an attendant cost or operability penalty.

5.3.4 Radiative and Unaccounted-For Losses

Radiative losses are a direct function of the surface emissivity, the fourth power of the absolute temperature, and the surface area. These radiative losses, as well as convective losses, occur from the boiler walls, steam pipes and other equipment where a temperature differential exists.

In estimating these losses, AFBC surface area plus piping was assumed equal to an equally rated conventionally-fired furnace plus piping. While early units had smaller total surface areas, increased freeboard in later designs has resulted in AFBC units with total volumes roughly equal to those of conventional units.[22]

A combined radiative and convective heat transfer coefficient of 15,560 J/m^2 - ^{o}C (2.5 Btu/ft^2 - ^{o}F) was determined from reference to Perry's Handbook of Chemical Engineering.[23] An average surface temperature of 200^{o}C (400^{o}F) was assumed. Dimensional proportions of equal height and depth, and width equal to one-half the height were used for heat loss calculations. The calculated losses include contingency losses such as piping, blowdown, and other small, intermittent losses. The resultant losses decrease from 3 percent down to 1.5 percent of the total heat input when the size is increased from 8.8 MW$_t$ up to 58.6 MW$_t$. Table 58 shows the losses in kW.

TABLE 58. RADIATIVE, CONVECTIVE, AND OTHER UNACCOUNTED LOSSES

Boiler capacity MW$_t$ (10^6 Btu/hr)	Loss by boiler type - KW
	Conventional/AFBC
8.8 (30)	265
22 (75)	479
44 (150)	750
58.6 (200)	903

These estimates reflect the economy-of-scale savings which result due to continuously decreasing surface-to-volume ratios with increasing boiler size.

5.3.5 Total Inherent Energy Penalties

All inherent losses associated with AFBC and uncontrolled conventional coal-fired industrial boilers (from Tables 55, 56, 57, and 58) are summed in Table 60 for each case - low, medium, and high reactivity sorbents; SIP, moderate, intermediate, and stringent control levels; and subbituminous, Eastern low sulfur, and Eastern high sulfur coals. There is no variability for the conventionally-fired boilers except by coal type and boiler capacity. Fuel-to-steam thermal efficiencies are estimated based on these inherent losses.

The FluiDyne unit reported in Table 59 is the 1 m × 1.62 m air heater with primary cyclone recycle. The B&W unit is a 3 ft × 3 ft test bed with no recycle capability. The Enköping unit is a 10 ft × 10 ft commercial steam generator capable of firing coal, oil, and gas.

TABLE 59. INHERENT LOSSES AS PERCENT OF THERMAL INPUT

Unit	GCA Estimate	FluiDyne [24]	B&W [25] 3 ft × 3 ft	Enköping [26]
Flue gas loss	10.5	22.7	22.7	5.6
(Flue gas losses - adjusted)*	-	(7.0)	(13.8)	(6.9)
Solids loss	1.6	1.9	0.6	1.5
Radiative loss	3.0	3.0	6.8	0.5
Combustion loss	3.0	1.7	17.4	0

*Losses in flue gas are adjusted to ITAR Design Conditions.

$$\text{Loss}_{\text{actual}} = \frac{\text{Combustion air}_{\text{design}}}{\text{Combustion air}_{\text{actual}}} \times \frac{\text{Temp}_{\text{design}}}{\text{Temp}_{\text{actual}}} \times \text{Loss}_{\text{design}}$$

TABLE 60. INHERENT ENERGY LOSSES OF UNCONTROLLED CONVENTIONAL BOILERS AND AFBC BY COAL SULFUR CONTENT, CONTROL LEVEL, AND SORBENT REACTIVITY - kW

				BOILER CAPACITY-MW							
				8.8		22		44		58.6	
COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	1830.	1628.	4391.	3885.	8524.	7562.	9286.	9986.
		LOW	4.2	1830.	1704.	4391.	4076.	8524.	7943.	9286.	10493.
		HIGH	2.3	1830.	1543.	4391.	3674.	8524.	7139.	9286.	9421.
	I 85%	AVERAGE	2.9	1830.	1601.	4391.	3820.	8524.	7431.	9286.	9811.
		LOW	3.8	1830.	1678.	4391.	4010.	8524.	7812.	9286.	10319.
		HIGH	2.1	1830.	1534.	4391.	3651.	8524.	7092.	9286.	9360.
	M 78.7%	AVERAGE	2.5	1830.	1577.	4391.	3759.	8524.	7310.	9286.	9650.
		LOW	3.4	1830.	1653.	4391.	3950.	8524.	7691.	9286.	10158.
		HIGH	1.8	1830.	1518.	4391.	3611.	8524.	7014.	9286.	9255.
	SIP 56%	AVERAGE	1.0	1830.	1485.	4391.	3529.	8524.	6850.	9286.	9036.
		LOW	1.2	1830.	1502.	4391.	3572.	8524.	6934.	9286.	9149.
		HIGH	0.8	1830.	1468.	4391.	3487.	8524.	6765.	9286.	8923.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	1608.	1445.	3835.	3428.	7434.	6648.	7843.	8767.
		LOW	3.7	1608.	1461.	3835.	3470.	7434.	6731.	7843.	8878.
		HIGH	2.0	1608.	1430.	3835.	3391.	7434.	6574.	7843.	8668.
	M 75%	AVERAGE	2.2	1608.	1437.	3835.	3408.	7434.	6607.	7843.	8713.
		LOW	3.2	1608.	1455.	3835.	3454.	7434.	6700.	7843.	8836.
		HIGH	1.6	1608.	1426.	3835.	3380.	7434.	6552.	7843.	8639.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	1814.	1635.	4351.	3904.	8463.	7600.	9798.	10037.
		LOW	3.6	1814.	1651.	4351.	3944.	8463.	7680.	9798.	10143.
		HIGH	2.0	1814.	1623.	4351.	3874.	8463.	7538.	9798.	9954.
	M 75%	AVERAGE	2.2	1814.	1629.	4351.	3889.	8463.	7564.	9798.	9945.
		LOW	3.2	1814.	1647.	4351.	3933.	8463.	7658.	9798.	10113.
		HIGH	1.6	1814.	1618.	4351.	3862.	8463.	7516.	9798.	9925.

Table 59 lists the relative inherent loss attributable to each identified component for AFBC and three operating units. The range in the flue gas losses for the operating units presented in Table 59 are a function of excess air rates, flue gas exhaust temperatures, and fuel analyses which differ from the ITAR design assumptions. The row titled (Flue Gas Losses - Adjusted) represents estimated losses at the three units after compensating for the differences in excess air and temperature. Any remaining differences are a function of fuel analysis and water content.

The rather high combustion loss of over 17 percent reported for the B&W 3 ft × 3 ft is not considered representative for fluidized beds. The 6 ft × 6 ft unit at B&W (which is an improved design) routinely achieves 91 to 96 percent combustion efficiency. The combustion efficiency reported for the FluiDyne unit is for coal combustion with primary cyclone recycle.

5.4 ENERGY IMPACT OF SO_2 CONTROL BY AFBC

The energy impact of SO_2 control is defined as the increase (or decrease) in energy requirements for the controlled FBC case, as compared to the conventionally-fired uncontrolled boiler.* Comparisons are made on the basis of total energy requirements; i.e., auxiliary losses plus inherent losses. For conventional SO_2 control methods, where some energy consuming device is added onto the conventional boiler, a net energy penalty must ensue. In the case of AFBC, the conventional boiler is eliminated and replaced by an integrated system of steam raising and SO_2 control within the same vessel. The net result is, in many instances, a net reduction in energy requirements, which in turn is reported as a negative energy penalty.

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*This reporting mode was developed to facilitate quantification of the energy penalty associated with implementing a specific technology as an SO_2 control option.

5.4.1 Efficiency

The efficiency is calculated on the basis of boiler input minus total losses.* Calculated efficiencies for AFBC are presented in Table C-17 and Figure 51. The conventionally-fired system efficiencies are included for comparative purposes.

Boiler efficiency improves with increasing boiler size, decreasing coal sulfur content, decreasing SO_2 control level, and increasing sorbent reactivity. The relative importance of these variables with respect to efficiency, in order of decreasing effect, is:

1. Coal sulfur;
2. Sorbent reactivity;
3. Boiler capacity; and
4. Control level.

While the ranking of these variables is somewhat a function of the assumptions incorporated within the analysis, the range considered is sufficiently broad that the results should be applicable to most commercial situations.

Efficiency estimates for the small boiler (8.8 MW_t) range from a low of 78.8 percent for Eastern high sulfur coal, stringent control, and low reactivity stone up to 82.2 percent for Eastern low sulfur coal, moderate control, and high reactivity sorbent. For the large (58.6 MW_t) boiler, efficiencies range

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*Total losses are auxiliary plus inherent losses.

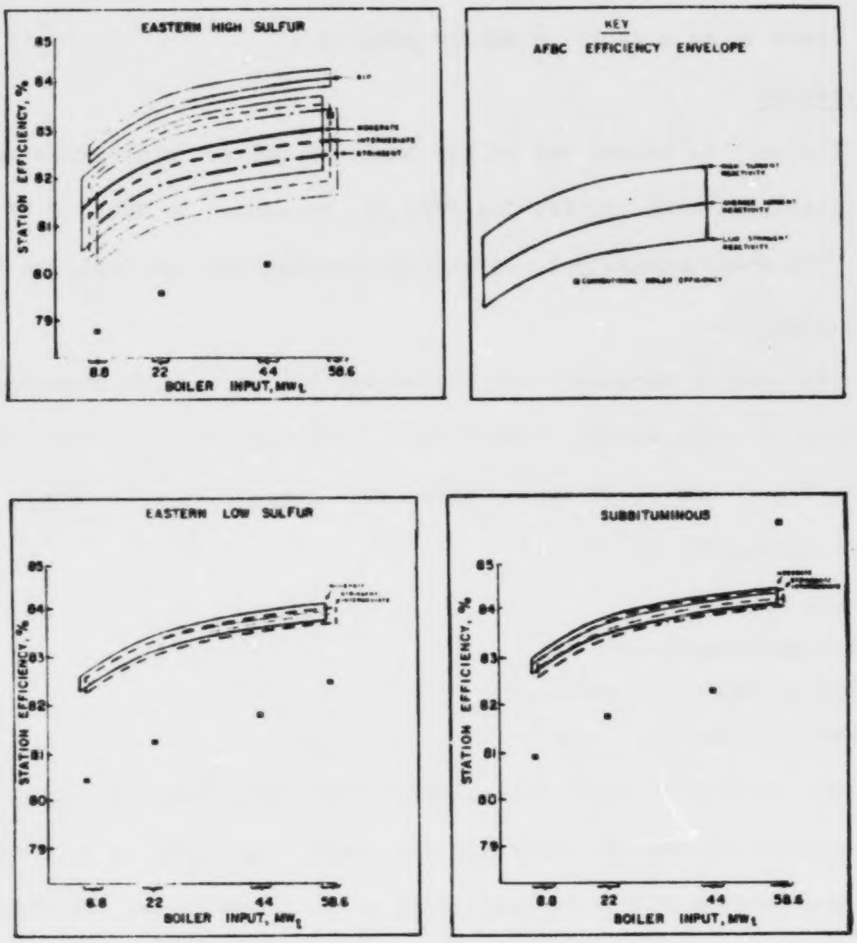

Figure 51. Station efficiency for AFBC and uncontrolled conventional boilers.

from a low of 80.3 percent for Eastern high sulfur coal, stringent control, and low sorbent reactivity up to 83.6 percent for Eastern low sulfur coal, moderate control, and high reactivity sorbent.

The efficiency of AFBC is as high or higher than the efficiency of a comparably-sized stoker-fired boiler for all cases. The pulverized coal-fired unit is more efficient than any AFBC option considered.

Overall fuel-to-steam* efficiencies reported in the literature for operating units are generally within the range covered by the GCA estimates. For example, CE[27] reported an efficiency of 81.8 percent and Johnston Boiler[28] reported an efficiency of 81.4 percent for high-sulfur high-ash coal, and 83.4 percent for low-sulfur low-ash coal. These efficiencies are all within the range of fuel-to-steam efficiencies estimated by GCA. The minimum efficiency estimated by GCA is 81.0 percent and the maximum is 83.8 percent for the 8.8 MW_t boiler.

5.4.2 Energy Penalty as kW/kg SO_2 Removed

Calculated energy differentials are divided by kilograms of SO_2 removed. This resultant value (kW/kg SO_2 removed) is the measure of effectiveness of an SO_2 control device for the technology (in this case, AFBC) in question. The kW/kg SO_2 removed calculated for each case under consideration is presented in Table C-18. Table 61 is an abbreviated listing of the energy penalty range expected for each coal type and boiler size over the range of control levels and sorbent reactivities investigated.

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*Fuel-to-steam efficiency excludes auxiliary losses.

TABLE 61. RANGE OF kW/kg SO_2 REMOVED BY COAL TYPE AND BOILER SIZE

	Boiler capacity - MW_t			
	8.8	22	44	58.6
Eastern high sulfur*	-12.4 to -1.0	-12.2 to -0.8	-11.7 to -0.5	1.3 to 7.6
Eastern low sulfur†	-16.2 to -9.5	-15.3 to -8.7	-14.2 to -7.7	27.7 to 34.9
Subbituminous†	-19.0 to -12.3	-18.1 to -11.5	-16.8 to -10.3	15.0 to 20.5

*Range is from SIP up to stringent SO_2 control for low to high reactivity sorbent.

†Range is from moderate up to stringent SO_2 control for low to high reactivity sorbent.

Examination of Table 61 reveals that the energy savings of AFBC over uncontrolled conventional units is greatest for the smaller units burning low sulfur coal. As unit size and/or coal sulfur increase, the energy savings for AFBC decrease. Finally, for the largest unit considered (58.6 MW_t), the uncontrolled conventionally-fired unit is more energy efficient than AFBC.

5.4.3 Efficiency of AFBC as a Percentage of Thermal Input

The energy impact of controlling SO_2 by AFBC and the increase in energy requirements when control levels more stringent than SIP are adopted are presented in Tables 62 and 63. The values of energy consumption are presented in terms of: energy consumed by control device; and percent change in energy use, compared to uncontrolled conventional boilers and AFBC boilers with SO_2 control at an average SIP level.

The impact of controlling SO_2 to an average SIP level of 1,075 ng/J (2.5 $lb/10^6$ Btu) is germane only when burning Eastern high sulfur coal where the required SO_2 reduction is 56 percent. The SIP control level does not apply to the Eastern low sulfur and subbituminous coals. The SIP energy requirements, as well as energy requirements for all other options considered, are presented in Table C-16.

These values were used as a basis to calculate entries in the last column of Tables 62 and 63. The incremental energy requirements between SIP control and the more stringent control levels ranges from 0.61 to 2.41 percent. The percent increase over uncontrolled conventional-firing ranges from -2.6 up to 3.2 for all cases considered. The reported percentage increase is calculated as follows:

TABLE 62. ENERGY CONSUMPTION FOR SO_2 CONTROL FOR AFBC COAL-FIRED BOILERS, 8.8 MW_t (30×10^6 Btu/hour) CAPACITY

System							Energy consumption		
Standard boiler									
Heat input									
MW_t	(MBtu/hr)	Fuel type	Level of control	SO_2 reduction %	Sorbent reactivity	Ca/S ratio	Energy consumed KW	Percent increase in energy use over uncontrolled conventional boiler as percent of boiler input	Percent change in energy use over SIP controlled AFBC boiler
8.8	(30)	Eastern high sulfur (3.5% S)	Stringent	90	Average	3.3	-117	-1.33	1.70
					Low	4.2	- 37	-0.42	2.42
					High	2.3	-205	-2.33	0.91
			Intermediate	85	Average	2.9	-144	-1.64	1.40
					Low	3.8	- 65	-0.74	2.10
					High	2.1	-215	-2.44	0.80
			Moderate	78.7	Average	2.5	-170	-1.93	1.10
					Low	3.4	- 91	-1.03	1.81
					High	1.8	-231	-2.63	0.61
			SIP	58.6	Average	1.0	-267	-3.03	
					Low	1.2	-250	-2.84	
					High	0.8	-285	-3.24	
		Eastern low sulfur (0.9% S)	Stringent or Intermediate	83.9	Average	2.8	- 88	-1.00	
					Low	3.7	- 71	-0.81	
					High	2.0	-103	-1.17	
			Moderate	75	Average	2.2	- 97	-1.10	
					Low	3.2	- 77	-0.88	
					High	1.6	-108	-1.23	
		Subbituminous (0.6% S)	Stringent or Intermediate	83.2	Average	2.7	-104	-1.18	
					Low	3.6	- 88	-1.00	
					High	2.0	-177	-2.01	
			Moderate	75	Average	2.2	-111	-1.26	
					Low	3.2	- 92	-1.05	
					High	1.6	-122	-1.39	

TABLE 63. ENERGY CONSUMPTION FOR SO_2 CONTROL FOR AFBC COAL-FIRED BOILERS, 58.6 MW_t (200×10^6 Btu/hr) CAPACITY

System							Energy consumption		
Standard boiler			Level of control	SO_2 reduction %	Sorbent reactivity	Ca/S ratio	Energy consumed KW	Percent increase in energy use over uncontrolled conventional boiler	Percent change in energy use over SIP controlled AFBC boiler
Heat input		Fuel type							
MW_t	(MBtu/hr)								
58.6	(200)	Eastern high sulfur (3.5% S)	Stringent	90	Average	3.3	1,327	2.26	1.72
					Low	4.2	1,855	3.17	2.42
					High	2.3	741	1.26	0.92
			Intermediate	85	Average	2.9	1,143	1.95	1.40
					Low	3.8	1,671	2.85	2.42
					High	2.1	674	1.15	0.80
			Moderate	78.7	Average	2.5	972	1.66	1.11
					Low	3.4	1,500	2.56	1.81
					High	1.8	561	0.96	0.61
			SIP	56	Average	1.0	321	0.54	
					Low	1.2	438	0.75	
					High	0.8	204	0.35	
		Eastern low sulfur (0.9% S)	Stringent or Intermediate	83.9	Average	2.8	1,479	2.52	
					Low	3.7	1,594	2.72	
					High	2.0	1,377	2.35	
			Moderate	75	Average	2.2	1,422	2.43	
					Low	3.2	1,550	2.65	
					High	1.6	1,345	2.30	
		Subbituminous (0.6% S)	Stringent or Intermediate	83.2	Average	2.7	794	1.35	
					Low	3.6	904	1.54	
					High	2.0	708	1.21	
			Moderate	75	Average	2.2	749	1.28	
					Low	3.2	872	1.49	
					High	1.6	676	1.15	

$$\% \text{ increase} = \frac{(\text{loss})_{\text{AFBC}} - (\text{loss})^{*}}{\text{Total thermal input}} \times 100$$

where * represents either uncontrolled conventional boiler loss or AFBC SIP-controlled loss. Although the energy envelopes overlap, the conclusions to be drawn are quite clear. For a given sorbent reactivity, SO_2 control level variability has a significant energy impact only for Eastern high sulfur coal. When highly reactive sorbents are used in the large boiler, all coals have nearly the same energy penalty (~1 percent). For the low reactivity sorbents, high sulfur Eastern coal usage is accompanied by an increase of 2.6 percent to 3.2 percent in the large boiler (58.6 MW_t) energy requirements. This range is a function of control level variability.

5.5 SENSITIVITY ANALYSIS

Several parameters which could be expected to affect the energy consumption of an AFBC system were varied through the extremes of a plausible range. The variables examined were excess air, calcium-to-sulfur ratio, combustion efficiency, sorbent reactivity, and spent solids heat recovery. A baseline around which these parameters were varied was also defined. The base conditions as well as the range of each parameter investigated are tabulated in Table 64. Boiler efficiency was selected to measure the effect of parametric variation on energy requirements. Boiler efficiency is defined as:

$$\text{efficiency} = ([\text{thermal input} - \text{inherent losses}]/\text{thermal input}) \times 100$$

The conventional boiler parameters were held constant throughout this analysis.

The results presented for each parameter are generated with a computerized mass and energy balance. For each set of conditions, all necessary parameters are fed into the program. A mass balance is then performed for the specified

TABLE 64. FBC PARAMETRIC CONSIDERATIONS (EASTERN HIGH SULFUR COAL)

Parameter	Std. Condition	Range
Excess air, %	20	0 - 100
Combustion efficiency, %	97	80 - 100
Ca/S ratio, m/m	3.5	1 - 10
SO_2 control efficiency (sorbent reactivity, %)	90	70 - 95
Coal Sulfur, %	3.5	1 - 10
Coal HHV, Btu/lb	11,800	--
Spent solids heat recovery, %	0	0 - 100
(Spent solids temp., ^{o}F)	1,500	1,550 - 300
Flue gas temperature, ^{o}F	350	--
Bottom Ash, %	90	--

conditions. The results of this mass balance are used to determine heat losses around the furnace. These losses are summed to arrive at a calculated boiler efficiency.

5.5.1 Calcium to Sulfur Ratio

The calculated effect of calcium-to-sulfur ratio on boiler efficiency is linear based on the results obtained when the Ca/S ratio is varied from 0 to 10. Because the effect is linear, an equation of the form

$$\text{Efficiency} = a\,(\text{Ca/S}) + b$$

was determined by linear least squares regression analysis for each boiler. The general equation, the conventional boiler efficiency, and the breakeven Ca/S ratio are presented in Table 65 and Figure 52.

TABLE 65. GENERAL EQUATIONS RELATING BOILER EFFICIENCY TO Ca/S FOR EASTERN HIGH SULFUR COAL

Boiler - MW_t	Equation	Conventional Boiler Efficiency	Breakeven Ca/S
8.8	$E^* = -0.963 \times (\text{Ca/S})^\dagger + 85.0$	79.5	5.73
22	$E = -0.963 \times (\text{Ca/S}) + 85.9$	80.4	5.66
44	$E = -0.963 \times (\text{Ca/S}) + 86.3$	81.0	5.53
58.6	$E = -0.963 \times (\text{Ca/S}) + 86.5$	84.2	2.38

[*] Boiler efficiency.

[†] Calcium-to-sulfur ratio.

This breakeven Ca/S is determined by substitution of the conventional boiler efficiency into the generalized equation. Any Ca/S requirement less than the breakeven point results in AFBC operation with a higher efficiency than the uncontrolled unit. The breakeven Ca/S ratio of 2.38 for the 58.6 MW_t unit indicates that, under the assumptions upon which this study is based, any lower Ca/S is sufficient for AFBC technology to exceed the efficiency of pulverized coal-fired technology.

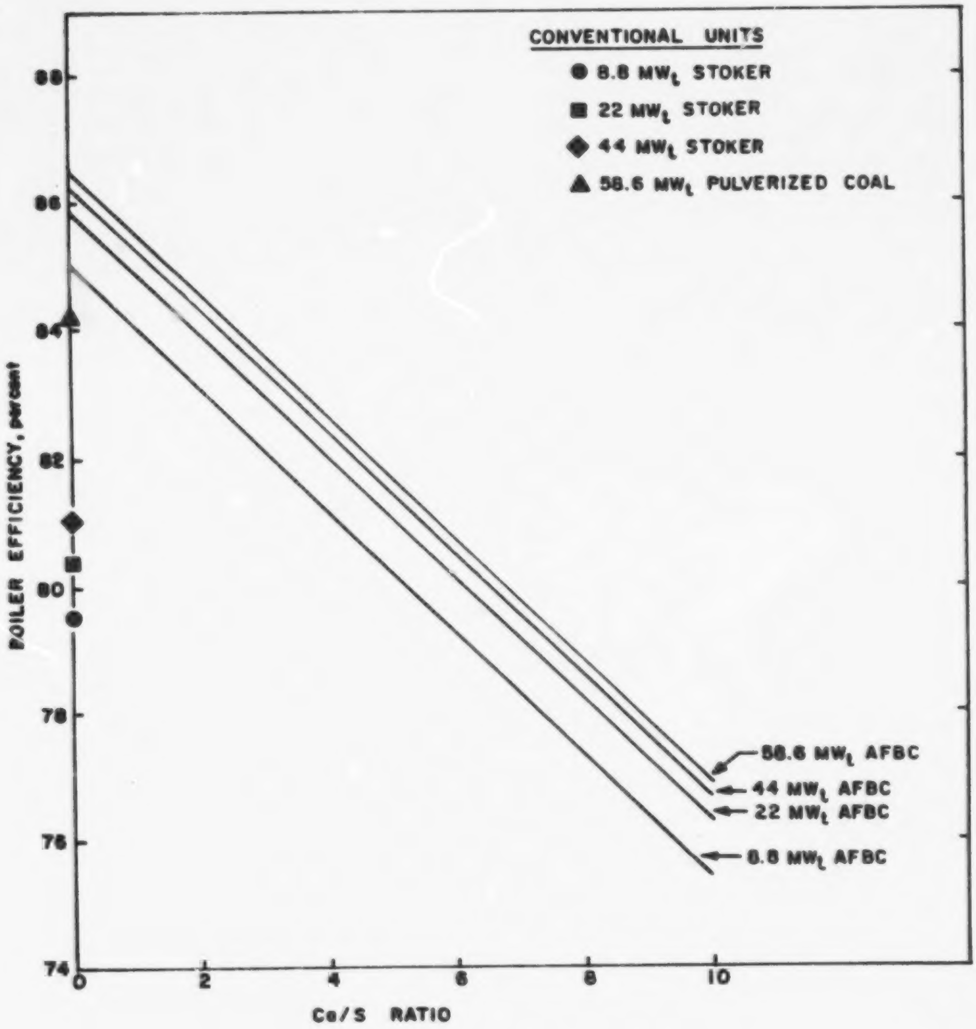

Figure 52. Boiler efficiency as a function of Ca/S molar feed ratio.

If AFBC Ca/S is maintained below 5.5 for the 8.8 - 44 MW_t units, AFBC is more efficient than conventional firing.

Calcium-to-sulfur ratio for a given system is a function of system design, sorbent reactivity, and sorbent particle size. System designs, incorporating increased bed depth or lower superficial velocity, such as proposed for best systems as opposed to commercially offered systems, can decrease sorbent requirements. Sorbent reactivity, which varies significantly among the sorbents tested, also affects sorbent requirements.

Implementation of any or all of these options (deeper beds, lower gas velocities, smaller sorbent particles, and more reactive sorbent) can increase boiler efficiency considerably. Each reduction of 1 in the Ca/S ratio improves boiler efficiency by 0.96 percent, as illustrated by Figure 52.

Considering the Foster-Wheeler Georgetown design with Greer limestone as an example (see Section 3, Table 21 where commercial and best systems are compared), Ca/S estimates are ~5.0 for a commercial system and 2.8 for best system conditions. This assumes stringent control and high sulfur coal as in the sensitivity assumptions (Appendix C, Table C-3). The estimated efficiency improvement in this example of using best system conditions is 2.1 percent.

5.5.2 Sorbent Reactivity

Recognizing that all sorbents are not equally capable of capturing SO_2 under identical conditions, the percent sulfur retained was varied while maintaining the Ca/S ratio constant. This analysis, as expected, indicates little overall effect on efficiency. As sorbent sulfur capture capability ranged from 70 percent up to 100 percent, boiler efficiency varied by roughly 0.5 percent.

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5.5.3 Spent Solids Heat Recovery

When spent solids are withdrawn from an FBC, sensible heat is lost with the solids. Some designs recover this sensible heat while others simply reject this heat as waste. To determine the effect on boiler efficiency of waste heat recovery, sensible heat recovery was varied from 0 up to 100 percent. Boiler efficiency increases by roughly 1 percent over the entire range from zero heat recovery to total heat recovery.

5.5.4 Coal Drying Requirement

Even though no coal drying is required for overbed coal feed AFBC systems, some commercially offered systems are designed for underbed feeding where coal drying is required. Because many commercial systems may require drying, an analysis of the effect on efficiency of coal moisture removal requirements was performed. As coal moisture varied, the coal analysis (and heating value) were normalized to compensate for the increased surface moisture.

Table 66 presents the linear equations relating boiler efficiency to coal drying requirements. In this analysis, spreader stoker-firing exhibits the least dependency on moisture content because no drying is required.

TABLE 66. RELATION BETWEEN BOILER EFFICIENCY AND COAL DRYING REQUIREMENTS

Boiler capacity MW_t	AFBC efficiency	Conventional efficiency	Breakeven moisture content
8.8	$-0.115P^* + 8.164$	$-0.102P + 79.56$	39.2
22	$-0.155P + 82.48$	$-0.102P + 80.40$	39.2
44	$-0.155P + 82.96$	$-0.102P + 80.98$	37.4
58.6	$-0.155P + 83.12$	$-0.155P + 84.73$	-

*P = percent moisture removed from coal.

The fluidized bed and pulverized coal-fired units exhibit identical dependency on coal moisture.

The breakeven moisture content is also listed in Table 66. At this moisture content the stoker and AFBC boiler efficiencies for a given boiler size are identical. The rather high breakeven points indicate that even should drying be required for AFBC, stokers will still be less energy efficient. The absence of a breakeven point for AFBC versus pulverized-firing results because both technologies are assumed to require the same percentage moisture removal.

No moisture content, under these design assumptions, is sufficiently low for AFBC-fired units to achieve higher efficiency than pulverized-fired units. For the smaller units (8.8 MW_t to 44 MW_t), any coal moisture removal requirement less than the breakeven point is sufficiently low for AFBC units to operate more efficiently than conventionally-fired stoker units.

5.5.5 Excess Air Effect

Excess air was calculated on the basis of Eastern high sulfur coal use with 97 percent combustion efficiency. Excess air is the percentage air introduced in excess of that required for stoichiometric combustion. The range examined is from 0 to 100 percent.

The effect of excess air variation is presented in Figure 53. As excess air increases, boiler efficiency decreases. The rate of decrease is slightly nonlinear. Each 10 percent increase in excess air is accompanied by roughly a 0.5 percent decrease in boiler efficiency.

The efficiencies of the conventional units are included for comparative purposes. To obtain efficiency equivalence between AFBC units and stokers,

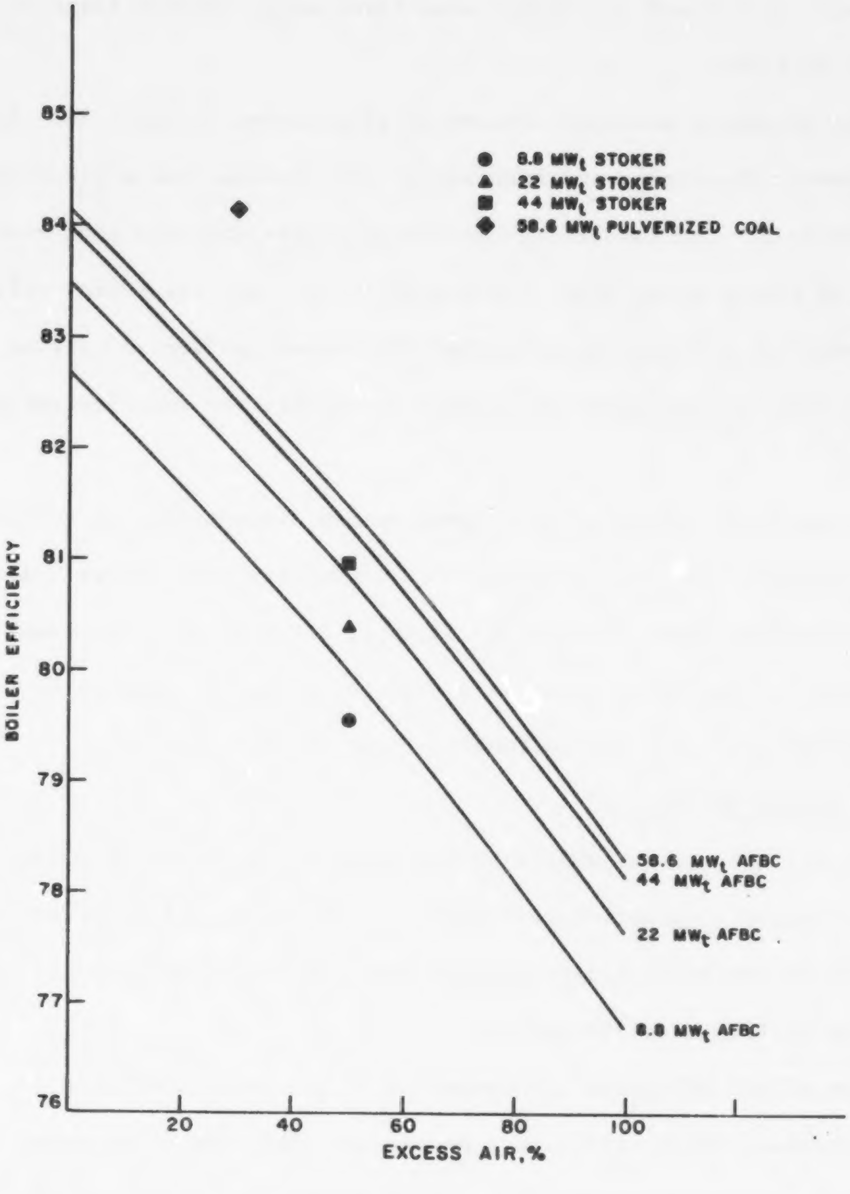

Figure 53. Boiler efficiency as a function of excess air rate.

AFBC units could run at excess air rates as high as 55 percent. An AFBC unit would need to operate with zero excess air to achieve the efficiency of a pulverized coal-fired unit.

5.5.6 Combustion Efficiency

The effect of combustion efficiency on boiler efficiency is linear. As in the case of Ca/S ratio, linear equations relating combustion efficiency to boiler efficiency were determined by regression analysis. These equations, along with the conventional technology boiler efficiencies, were then used to determine the combustion efficiency necessary for equivalent boiler efficiencies for the two technologies. The general equations, the conventional system efficiencies, and the breakeven combustion efficiencies are presented in Table 67.

TABLE 67. GENERAL EQUATION RELATING BOILER EFFICIENCY TO COMBUSTION EFFICIENCY

Boiler - MW_t	Equation	Conventional boiler efficiency	Breakeven combustion efficiency
8.8	$E^{*} = 0.891 \times (CE)^{\dagger} - 4.837$	79.5	94.6
22	$E = 0.891 \times (CE) - 3.95$	80.4	94.7
44	$E = 0.891 \times (CE) - 3.56$	81.0	94.8
58.6	$E = 0.891 \times (CE) - 3.314$	84.2	98.2

[*] Boiler efficiency.

[†] Combustion efficiency.

In all cases, sufficiently high combustion efficiency will result in AFBC boiler efficiency as good as or better than conventional boiler technology. The ability of AFBC technology to achieve these combustion efficiencies has not yet been demonstrated.

5.6 ENERGY IMPACT OF NO_x CONTROL

As discussed in Chapter 3, commercial-scale AFBC units should generally be able to achieve all three levels of NO_x control without major adjustments to design/operating conditions. Thus, the desired levels of NO_x control should be achievable with no additional energy impact on the AFBC system.

5.7 ENERGY IMPACT OF PARTICULATE CONTROL

Energy required for final particulate control in AFBC industrial boilers is expected to be similar to that resulting from application of conventional particle control devices on conventional boilers. Particulate emissions from a conventional boiler are ash and char. The emissions from an AFBC are limestone, spent bed material, ash, and char. At 177°C (350°F), AFBC flue gas rates are less than the values noted for the four conventional coal-fired boilers. The difference is due to the difference in excess air values. The conventional coal-fired boilers operate at excess air rates between 30 and 50 percent, while the AFBC boilers operate at 20 percent excess air. On this basis, it may be projected that the requirements for particulate control in conventional systems provide a conservative indication of energy impact associated with final particulate control operation in AFBC industrial boilers.

Table 68 presents a summary of energy requirements for final particulate control for coal-fired AFBC industrial boilers. For each level of control, energy use is shown for the systems discussed in Section 3.0. Estimates of energy losses/auxiliary requirements in an uncontrolled conventional boiler were obtained from "Technology Assessment Report for Industrial Boilers: Particulate Control."[29]

TABLE 68. ENERGY CONSUMPTION FOR BEST PARTICULATE CONTROL COAL-FIRED AFBC BOILERS

System				Energy consumption		
Standard boiler Heat input MW	(MBtu/hr)	Type & level of control	Control efficiency %	Range in energy consumed by control device KW[*]	Percent increase in energy[†] use over uncontrolled conventional boiler	Percent change in energy use over SIP controlled AFBC boiler
8.8	(30)	Stringent				
		FF	94.0 - 98.7	15.6 - 16.4	0.89 - 0.96	0
		ESP	97.6 - 97.9	13.0 - 16.0	0.74 - 0.91	0.59 - 0.72
		Intermediate				
		FF	80.0 - 98	15.6 - 16.4	0.89 - 0.93	0
		ESP	92.0 - 92.9	9.3 - 11.4	0.53 - 0.65	0.35 - 0.42
		MC	80 - 82.1	15.6 - 16.0	0.89 - 0.91	0
		Moderate				
		FF	50 - 95	15.6 - 16.4	0.89 - 0.93	0
		ESP	80 - 82.1	6.5 - 8.2	0.37 - 0.47	0.16 - 0.21
		MC	50 - 82.1	15.6 - 16.0	0.89 - 0.91	0
		SIP				
		FF	≤88	15.6 - 16.4	0.89 - 0.93	-
		ESP	52 - 57.1	4.0 - 5.0	0.23 - 0.29	-
		MC	≤82	15.6 - 16.0	0.89 - 0.91	-
22	(75)	Stringent				
		FF	94 - 98.7	38.4 - 41.2	0.93 - 1.00	0
		ESP	97.6 - 97.9	32.8 - 40.9	0.80 - 0.99	0.65 - 0.81
		Intermediate				
		FF	80 - 98	38.4 - 41.2	0.93 - 1.00	0
		ESP	92 - 92.9	23.5 - 29.5	0.57 - 0.72	0.38 - 0.48
		MC	80 - 82.3	38.4 - 40.0	0.93 - 0.97	0
		Moderate				
		FF	50 - 95	38.4 - 4.12	0.93 - 1.00	0
		ESP	80 - 82.3	16.6 - 20.9	0.40 - 0.51	0.19 - 0.23
		MC	50 - 82.3	38.4 - 40.0	0.93 - 0.97	0
		SIP				
		FF	≤88	38.4 - 41.2	0.93 - 1.00	-
		ESP	52 - 57.5	10.1 - 12.7	0.25 - 0.31	-
		MC	≤82	38.4 - 40.0	0.93 - 0.97	-
44	(150)	Stringent				
		FF	94 - 99.5	77.6 - 82.6	0.98 - 1.04	0
		ESP	99.1 - 99.2	82.4 - 102.1	1.04 - 1.29	0.70 - 0.86
		Intermediate				
		FF	80 - 98	77.6 - 82.6	0.98 - 1.04	0
		ESP	96.9 - 97.3	63.2 - 78.6	0.80 - 0.99	0.41 - 0.51
		MC	80 - 82	77.6 - 80	0.98 - 1.01	0
		Moderate				
		FF	50 - 95	77.6 - 82.6	0.98 - 1.04	0
		ESP	92.3 - 93.2	48.8 - 60.6	0.61 - 0.76	0.20 - 0.25
		MC	50 - 80	77.6 - 80	0.98 - 1.01	0
		SIP				
		FF	≤88	77.6 - 82.6	0.98 - 1.04	-
		ESP	81.5 - 83.6	35.3 - 43.9	0.44 - 0.55	-
		MC	≤82	77.6 - 80	0.98 - 1.01	-
58.6	(200)	Stringent				
		FF	94 - 99.4	90.2 - 95.4	1.07 - 1.13	0
		ESP	99.2 - 99.3	99.3 - 124.0	1.18 - 1.47	0.63 - 0.77
		Intermediate				
		FF	80 - 98	90.2 - 95.4	1.07 - 1.13	0
		ESP	97.5 - 97.8	77.2 - 96.5	0.91 - 1.14	0.37 - 0.46
		MC	80 - 82	90.2 - 92.5	1.07 - 1.10	0
		Moderate				
		FF	50 - 95	90.2 - 95.4	1.07 - 1.13	0
		ESP	93.8 - 94.4	60.5 - 75.5	0.72 - 0.89	0.18 - 0.22
		MC	50 - 80	90.2 - 92.5	1.07 - 1.10	0
		SIP				
		FF	≤88	90.2 - 95.4	1.07 - 1.13	-
		ESP	85 - 86.7	44.4 - 55.8	0.53 - 0.66	-
		MC	≤82	90.2 - 92.5	1.07 - 1.10	-

[*] The energy consumed by the particle control device on an AFBC was assumed to be the same as its energy consumption on conventional boilers, taken from Reference 23.

[†] Energy refers to auxiliary plus inherent energy requirements.

Electrostatic precipitation energy use estimates for low sulfur coal combustion in conventional units were considered comparable to anticipated AFBC requirements where effective ESP operation would probably require hot side installation.

As in the case of AFBC SO_2 control, the energy impact of particulate control devices applied to AFBC is expressed in terms of the percentage increase in energy usage over that in an uncontrolled conventional boiler. Of course, a similar increase in energy usage would be experienced in a conventional unit.

The percent increase in energy use presented in Table 68 is calculated as follows:

Uncontrolled Conventional Boilers

$$\%\ \text{increase} = \frac{\text{Energy consumed by control device}}{\text{Total system energy requirements for uncontrolled conventional boiler}} \times 100$$

SIP-controlled AFBC boiler

$$\%\ \text{increase} = \frac{(\text{Energy consumed by control device}) - (\text{Energy consumed by control device for SIP control})}{(\text{Total system energy requirements for AFBC}) + (\text{Energy consumed by control device for SIP control})} \times 100$$

Energy use calculated on this basis associated with the full range of anticipated efficiency requirements; i.e., from 50 to 99.4 percent, is shown for fabric filters and multitube cyclones. This range is also covered for ESP's but in discrete steps. Interpolation of the data is necessary in cases where the specific control level of interest was not considered in the particulate control ITAR. Enough information is shown to indicate the relative differences in energy requirements using different control devices to support stringent, intermediate, and moderate particulate reduction levels.

5.7.1 Comparison of Fabric Filters and Electrostatic Precipitators

Fabric filters and ESP's were recommended for stringent control in Section 3.0, Table 26. Energy requirements at the stringent control level are similar for these two control methods. Fabric filters appear to have a slight advantage for the two larger boilers where removal requirements exceed 98 percent, whereas, ESP's have a slight advantage at intermediate and moderate control levels. (This effect may be a result of an assumption of constant pressure drop for fabric filters, regardless of control level.)

For control with fabric filters, the energy penalty ranges from between 0.89 to 0.96 percent for the 8.8 MW_t boiler up to 1.07 percent to 1.13 percent for the largest boiler. Because of the constant pressure drop assumption, there is no variation in energy penalty with control level for fabric filters. For stringent control with ESP technology, the 8.8 MW_t boiler penalty range is 0.74 to 0.91 percent and for the large boiler the range is 1.18 to 1.47 percent. For less stringent control this energy penalty is lower.

5.7.2 Impact of Multitube Cyclone Use

The energy penalty accompanying particulate control by multitube cyclone is only slightly less than for fabric filters. The range for cyclones is 0.89 to 0.93 percent for the 8.8 MW_t boiler and 1.07 to 1.10 percent for the 58.6 MW_t boiler.

When comparing SIP control level with the moderate, intermediate, and stringent levels under consideration, there is no associated energy penalty for the fabric filters or for the multitube cyclones (as a result of the constant pressure drop assumption). The SIP control energy difference is only a factor for ESP technology. When the moderate, intermediate, and stringent levels are compared to SIP control (see Table 68), the energy penalties are roughly as follows:

- moderate - 0.20 percent
- intermediate - 0.4 percent
- stringent - 0.65 percent

The effect of boiler size on energy penalty is miniscule.

It is thus projected that the optional particulate control levels can be supported by AFBC with conventional add-on particulate controls, with an attendant energy penalty of from 0.4 up to 1.15 percent, compared to a conventional uncontrolled boiler. The exact energy penalty is a function of control level, control device, and boiler size. Since particulate emissions (downstream of the primary cyclone) are a function of SO_2 control level, sorbent particle size, and primary cyclone efficiency, final particle control energy use is also a function of these factors, especially in the case of ESP control. ESP performance must be confirmed on the basis of sorbent resistivity and total sorbent loadings to determine above 95 percent are routinely achievable. Because of these unknowns, the energy estimates for FF and MC control have a higher confidence level than those noted for ESP operation. In conclusion, the energy impact of ESP control is a function of the SO_2 removal system but FF and MC energy use is not expected to be as sensitive to SO_2 control methodology, provided the constant pressure drop assumption is valid.

5.8 SUMMARY

5.8.1 SO_2 Control

The estimated energy requirements for SO_2 control when AFBC is compared to uncontrolled conventional systems ranges from -2.6 percent of thermal input to the boiler (which represents an energy savings) up to 3.2 percent (which represents an energy penalty). The wide range is principally a function of boiler size. Other variables which affect energy requirements are coal type and sorbent reactivity.

The level of SO_2 control in AFBC has a minor effect on the energy impact of the total system. This is illustrated in Table 69 which shows the differential changes in boiler efficiency as FBC design/operating parameters are varied through the full range considered in this report.

With Eastern high sulfur coal, boiler efficiency decreases by about 0.6 percent when control level is increased from moderate to stringent. This is the minimum differential for the parameters considered. The coal sulfur content proved to have the most significant effect on boiler efficiency.

TABLE 69. DIFFERENTIAL CHANGES IN BOILER EFFICIENCY VERSUS RANGE OF FBC DESIGN/OPERATING PARAMETERS

FBC design/operating parameter and range	Differential change in boiler efficiency
Sorbent reactivity - low to high*	1.83
Coal sulfur content - 0.6 to 3.5†	2.17
Boiler capacity - 8.8 to 58.6 MW_t†	1.47
SO_2 control level - moderate to stringent‡	0.58

*Stringent control, Eastern high sulfur coal.

†Stringent control, average sorbent reactivity.

‡Eastern high sulfur coal, average sorbent reactivity.

The comparison of AFBC and uncontrolled conventional boilers showed that for any of the three smaller boilers (8.8, 22, and 44 MW_t), AFBC boiler efficiency was 1 to 3 percent higher than conventional boiler efficiency considering all optional control levels and coal types. For the large boiler (58.6 MW_t), AFBC boiler efficiency was 1 to 3 percent lower than the conventional pulverized coal unit.

Of the total system losses, roughly 10 percent are auxiliary losses and 90 percent are inherent losses for the options investigated (see Table 70). The major component of the auxiliary losses is fan power. Fan power requirements comprise approximately two-thirds of the auxiliary power required in an FBC system. The principal inherent loss component, flue gas sensible heat loss accounts for roughly two-thirds of the inherent losses. Even the largest auxiliary component (fan power) is not particularly significant when total system losses are considered.

TABLE 70. TOTAL SYSTEM LOSSES RESULTING FROM EACH ENERGY COMPONENT CONSIDERED

Component	Uncontrolled Conventional KW	Uncontrolled Conventional Percent	AFBC KW	AFBC Percent
Auxiliary				
Coal Handling	6-35	0.3	6-35	0.3
Fan Power	42-227	2.3	115-766	6.4
Solids Handling	3-19	0.2	3-128	0.9
Pumping	18-125	1.2	18-125	1.0
Inherent				
Flue Gas	1065-7381	71.8	881-7170	59.1
Solids	13-72	0.7	1-142	10.3
Combustion	264-1318	13.4	264-1757	14.6
Radiative	265-903	9.9	265-903	8.4

Because flue gas desulfurization is the only widely commercialized sulfur emission control method for coal-fired steam raising, percentage energy requirements for the four most widely accepted systems are presented in Table 71, along with estimates of AFBC energy requirements. Flue gas desulfurization energy requirements vary as a function of coal sulfur level, SO_2 control level, and to a smaller extent, plant size.[30] Industrial fluidized-bed combustion energy requirements vary with coal sulfur level, sorbent reactivity, sulfur emission control level, and plant size.

TABLE 71. RANGE OF FGD[31] AND FBC PROCESS ENERGY REQUIREMENTS

SO_2 control method	Energy requirement (percent) increase over uncontrolled conventional boiler)
Lime/Limestone	2.6 to 3.7
Double Alkali	2.0 to 2.4
Sodium Scrubbing	2.0 to 2.6
Wellman-Lord	3.2 to 8.0
AFBC	-2.6 to 3.2

While the range presented for AFBC encompasses both double alkali and sodium scrubbing, the actual energy requirements would probably be lower than those estimated because the upper and lower limits of the range are mainly a function of sorbent reactivity. If only average sorbent reactivity is considered, the range is from -1.9 percent up to 2.5 percent. The negative value (-1.9 percent) indicates that AFBC system losses are less than uncontrolled conventionally-fired systems.

5.8.2 Particulate Control

Particulate control energy requirements range from 0.4 to 1.45 percent of total operating energy requirement if ESP's are used, and from 0.90 to 1.15 percent if fabric filters or multitube cyclones are used. ESP energy is a strong function of control efficiency, but FF and MC energy use is fairly independent of control efficiency.

5.8.3 NO_x Control

NO_x reduction to stringent, intermediate, or moderate levels can be achieved at standard FBC operating conditions, so that no auxiliary energy requirements are expected.

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6.0 FLUIDIZED-BED COMBUSTION ENVIRONMENTAL IMPACT

6.1 INTRODUCTION

This section provides an assessment of the environmental impact of adopting the "best systems" for emission control in atmospheric fluidized-bed combustion as applied to industrial-sized boilers.

In fluidized-bed combustion, the most prominent environmental impact is solid waste disposal. The "best system" design for FBC is based on minimizing the Ca/S ratio, and thus the amount of sorbent and solid waste which is necessary to achieve a given level of SO_2 reduction. Therefore, as "commercially offered" design/operating conditions approach "best system" conditions, the environmental impact will be reduced. The impact of SO_2 emissions will remain the same because specific SO_2 control levels are the frame of reference. The effect on NO_x and particulate emissions is uncertain, although NO_x may be reduced due to extended gas phase residence times.

6.1.1 Emission Streams

Figure 54 is a diagram showing the waste streams from a simplified FBC system. The pollutants from the system can be divided into the following categories:

- Stack gas - SO_2, NO_x and particulate emissions are the primary pollutants emitted in the stack gas. CO, hydrocarbons, and volatile trace element emissions may also be of concern. These latter mentioned pollutants are emitted at the same low level from FBC as from conventional coal-fired combustors. The environmental impact of these emissions is under continued investigation.

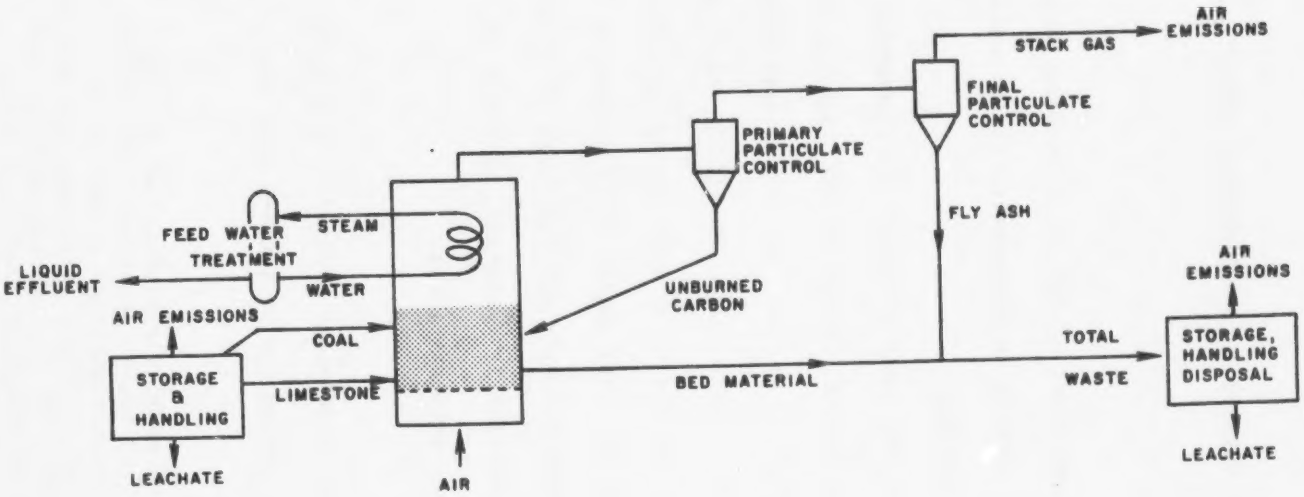
AIR EMISSIONS
STACK GAS
FINAL PARTICULATE CONTROL
PRIMARY PARTICULATE CONTROL
FLY ASH
FEED WATER TREATMENT
STEAM
LIQUID EFFLUENT
WATER
UNBURNED CARBON
AIR EMISSIONS
AIR EMISSIONS
COAL
TOTAL
STORAGE, HANDLING DISPOSAL
STORAGE & HANDLING
LIMESTONE
BED MATERIAL
WASTE
LEACHATE
AIR
LEACHATE

Figure 54. FBC flow diagram.

- Solid residue - Spent bed material and fly ash are the two types of solid residue produced by FBC systems. This residue creates the major environmental impact of FBC. CAO, $CaSO_4$, ash and impurities make up the solid residue. Handling and disposal problems arise from the potential heat release of the material upon contact with water, the high pH and high total dissolved solids attributed primarily to the high CaO content of the solid. The waste may also contain toxic trace elements, from the coal ash and limestone impurities which may be leachable. Care must be taken in designing handling and disposal systems, but based on current information, there is no reason to assume that waste disposal cannot be accomplished in an environmentally acceptable manner.

- Fugitive emissions - Coal, limestone and solid waste storage, handling, and onsite transportation may produce fugitive dust emissions and possibly even some low level radiation. These emissions are expected to be equivalent to those produced at the site of a conventional coal-fired facility with lime/limestone flue gas desulfurization.

- Water - Most effluents from an FBC plant are expected to be the same as those from conventional systems. The steam cycle discharges result from feed water treatment and boiler blowdown. These discharges should be equivalent to those produced in conventional boilers because FBC boiler designs are expected to follow existing boiler codes. Wastes from fireside boiler cleaning could differ from conventional systems, but such cleanings occur infrequently and should pose a minor impact. Water pollution also results from rainwater percolation through storage piles forming a leachate. Leachate from coal storage piles will be the same as that encountered in conventional systems.

6.1.2 Major Issues

Based on the data which are presently available, conclusions have been drawn by several investigators[1-4] that fluidized-bed combustion is an environmentally sound technology and no insurmountable pollution problems are foreseen. Further investigation, however, is recommended and is presently being undertaken on larger scale tests for air emissions, as well as solid waste disposal, including analysis of a wide range of possible pollutants not previously considered. The major issue of concern with respect to the environmental impact of FBC is the amount of the solid waste and disposal requirements.

The amount of spent solids produced at an industrial FBC boiler facility is primarily dependent upon: the unit capacity, the coal sulfur content and the level of SO_2 control desired. The approximate range of solid waste predicted under the size, fuel and control level guidelines of this study is 100 kg/hr (220 lb/hr) to nearly 4,000 kg/hr (8,800 lb/hr). Handling and disposal options need to be identified and studied because of the heat release properties of the material and the high pH and total dissolved solids.

6.1.2.1 Influence of RCRA--

The disposal options must take into account the states of FBC waste under the Resource Conservation and Recovery Act (RCRA, PL 94-580) as well as leachate characteristics affected by the National Pollution Discharge Elimination System (NPDES) and any other legislation governing the quality of the nation's waters. According to recent tests sponsored by EPA and carried out by Westinghouse Research Laboratories, FBC residues do not appear to be "hazardous" according to the procedures currently proposed under Section 3001 of RCRA. Using the Extraction Procedure proposed in the Federal Register,[5] tests showed that none of the eight species called out in the Federal Register exceeded the threshold of 10 times the national interim primary drinking water standards. The other criteria in Section 3001 are "ignitable," "reactive," and "corrosive" and they do not seem to apply to FBC waste, although no formal ruling has been made. The latter criteria could conceivably apply to FBC waste, but for the time being, current interpretations are that this applies to liquid wastes and not solids or leachates from solids.

The designation of FBC waste under the RCRA waste categories will continue to be an active research and regulatory issue for the near future. Solid wastes from electric utilities have been placed in a special high volume category; this seems a likely interim category for FBC waste until more data become available.

The process variables which have the greatest effect on the environmental impact of FBC solid waste (both quantity and composition) are those which determine the amount of sorbent used. They include the level of SO_2 emission control desired, the Ca/S molar feed ratio necessary to meet that level and the operating variables and conditions which are used to minimize the Ca/S ratio; i.e., sorbent particle size, gas phase residence time, sorbent reactivity, and bed temperature. The quantity of sorbent used affects most of the pollution emissions in some manner.

6.1.2.2 Multimedia Impact--

When analyzing the environmental impact of a given system and the control regulations applying to it, it is of the utmost importance to consider cross pollutant and multimedia effects; i.e., what is the impact of reducing one pollutant on the emission of the others. The other pollutants can be affected in two ways: (1) directly, producing a new or increased amount of byproduct, such as collected fly ash resulting from flue gas particulate removal; and (2) changing the conditions of the system such that they affect other pollutants from the system, such as increasing gas residence time to increase SO_2 capture, with the result of decreased NO emissions as an additional benefit. In this assessment of the environmental impact of SO_2, NO and particulate control on FBC, a multipollutant approach has been taken.

Other issues of environmental and commercial concern also need to be further investigated. SO_2 control performance of AFBC must be more fully demonstrated at the 0.67 second gas phase residence time and 500 μm average bed particle size which have been chosen to represent the "best" system. Particle control devices must be adequately demonstrated as applied to AFBC in order to prove applicability and reliability of these systems. Another issue that must be further investigated is the emission of trace elements from FBC industrial boilers.

Section 6.2 also includes a brief discussion of FBC versus conventional combustion with flue gas desulfurization as reported by several investigators. Although this is not in the scope of the project, it is felt that this comparison will give the reader a better perspective on the environmental impact of FBC compared to other coal-based combustion systems.

6.2 ENVIRONMENTAL IMPACT OF COAL-FIRED AFBC

The air pollution impact of AFBC industrial boilers will most likely be the same for "commercially offered" units as for the proposed "best system" of emission control, if the same levels of emission reduction are considered for each system. The discussion which follows, therefore, applies to both systems.

The solid waste impact, however, will vary between the systems, due to the variations in operating parameters between "commercially offered" systems and the "best" system, and the resultant differences in sorbent requirements to achieve equivalent levels of SO_2 reduction.

6.2.1 Air Pollution

6.2.1.1 SO_2 Emissions--

Tables 72a through 72d illustrate the SO_2 emissions from coal-fired atmospheric fluidized bed combustion boilers under varying conditions, which include four boiler capacities, three different coals, and three SO_2 control levels. The Ca/S ratios indicated in Table 72 are projected for AFBC design and operating conditions representing the "best" system for SO_2 control in AFBC with a sorbent of average reactivity (see Section 3.0).

TABLE 72. AIR POLLUTION IMPACTS FROM "BEST" AND "COMMERCIALLY OFFERED" SO_2 CONTROL SYSTEMS FOR COAL-FIRED FBC BOILERS (8.8 MW_t or 30 × 10^6 Btu/hr heat input)

System							Air Emissions						Secondary pollutants		
	Heat value				Type of control		SO_2				Other pollutants		Solid waste[#]		
%S	kJ/kg	(Btu/lb)	Control[†] level	Percent[‡] reduction	System	Ca/S	g/s	(lb/h)	ng/J	(lb/10^6 Btu)	Pollutant	Effect[§]	"Best" system kg/h	"Best" system (lb/h)	Commercial system kg/h (lb/h)
3.5	27,450	(11,800)	none	0	AFBC	0	21	(169)	2,424	(5.64)	NA	NA	NA		NA
0.9	32,100	(13,800)	none	0	AFBC	0	4.7	(37)	533	(1.24)	NA	NA	NA		NA
0.6	22,330	(9,600)	none	0	AFBC	0	4.5	(36)	512	(1.19)	NA	NA	NA		NA
3.5	27,450	(11,800)	M	78.7	AFBC	2.5	4.5	(36)	516	(1.20)	particulate	+	420	(925)	
3.5	27,450	(11,800)	I	85	AFBC	2.9	3.3	(26)	364	(0.85)	particulate	+	461	(1,016)	
3.5	27,450	(11,900)	S	90	AFBC	3.3	2.1	(17)	242	(0.56)	particulate	+	502	(1,105)	
0.9	32,100	(13,800)	M	75	AFBC	2.2	1.2	(9.3)	133	(0.31)	particulate	+	128	(281)	
0.9	32,100	(13,800)	I & S	83.9	AFBC	2.8	0.76	(6.0)	86	(0.20)	particulate	+	142	(310)	
0.8	32,100	(13,800)	S+	90	AFBC	3.3	0.45	(3.6)	52	(0.12)	particulate	+	152	(335)	
0.6	22,330	(9,600)	M	75	AFBC	2.2	1.1	(9.0)	128	(0.30)	particulate	+	133	(293)	
0.6	22,330	(9,600)	I & S	83.2	AFBC	2.7	0.76	(6.0)	86	(0.20)	particulate	+	144	(318)	
0.6	22,330	(9,600)	S+	90	AFBC	3.3	0.45	(3.6)	52	(0.12)	particulate	+	157	(345)	

TABLE 72b. AIR POLLUTION IMPACTS FROM "BEST" AND "COMMERCIALLY OFFERED" SO_2 CONTROL SYSTEMS FOR COAL-FIRED FBC BOILERS (22 MW_t or 75 × 10^6 Btu/hr heat input)

System							Air emissions						Secondary pollutants		
	Heat value				Type of control		SO_2				Other pollutants		Solid waste[#]		
%S	kJ/kg	(Btu/lb)	Control[†] level	Percent[‡] reduction	System	Ca/S	g/s	(lb/h)	ng/J	(lb/10^6 Btu)	Pollutant	Effect[§]	"Best" system kg/h	(lb/h)	Commercial system kg/h (lb/h)
3.5	27,450	(11,800)	none	0	AFBC	0	53	(423)	2,425	(5.64)	NA	NA	NA		NA
0.9	32,100	(13,800)	none	0	AFBC	0	12	(93)	533	(1.24)	NA	NA	NA		NA
0.6	22,330	(9,600)	none	0	AFBC	0	11	(89)	512	(1.19)	NA	NA	NA		NA
3.5	27,450	(11,800)	M	78.7	AFBC	2.5	11	(90)	516	(1.20)	particulate	+	1,052	(2,318)	
3.5	27,450	(11,800)	I	85	AFBC	2.9	8.0	(64)	364	(0.85)	particulate	+	1,155	(2,545)	
3.5	27,450	(11,800)	S	90	AFBC	3.3	5.3	(42)	242	(0.56)	particulate	+	1,255	(2,766)	
0.9	32,100	(13,800)	M	75	AFBC	2.2	2.9	(23)	133	(0.31)	particulate	+	314	(699)	
0.9	32,100	(13,800)	I & S	83.9	AFBC	2.8	1.9	(15)	86	(0.20)	particulate	+	352	(775)	
0.9	32,100	(13,800)	S+	90	AFBC	3.3	1.1	(9)	52	(0.12)	particulate	+	379	(835)	
0.6	22,330	(9,600)	M	75	AFBC	2.2	2.8	(22)	128	(0.30)	particulate	+	332	(734)	
0.6	22,330	(9,600)	I & S	83.2	AFBC	2.7	1.9	(15)	86	(0.20)	particulate	+	360	(794)	
0.6	22,330	(9,600)	S+	90	AFBC	3.3	1.1	(9)	52	(0.12)	particulate	+	392	(864)	

TABLE 72c. AIR POLLUTION IMPACTS FROM "BEST" AND "COMMERCIALLY OFFERED" SO_2 CONTROL SYSTEMS FOR COAL-FIRED FBC BOILERS (44 MW_t or 150 × 10^6 Btu/hr heat input)

System							Air emissions						Secondary pollutants		
													Solid waste[#]		
	Heat value				Type of control		SO_2				Other pollutants				
%S	kJ/kg	(Btu/lb)	Control[†] level	Percent[‡] reduction	System	Ca/S	g/s	(lb/h)	ng/J	(lb/10^6 Btu)	Pollutant	Effect[§]	"Best" system kg/h	(lb/h)	Commercial system kg/h (lb/h)
3.5	27,450	(11,800)	none	0	AFBC	0	107	(846)	2,425	(5.64)	NA	NA	NA		NA
0.9	32,100	(13,800)	none	0	AFBC	0	23	(186)	533	(1.24)	NA	NA	NA		NA
0.6	22,330	(9,600)	none	0	AFBC	0	23	(179)	512	(1.19)	NA	NA	NA		NA
3.5	27,450	(11,800)	M	78.7	AFBC	2.5	23	(180)	516	(1.20)	particulate	+	2,102	(4,635)	
3.5	27,450	(11,800)	I	85	AFBC	2.9	16	(128)	364	(0.85)	particulate	+	2,309	(5,089)	
3.5	27,450	(11,800)	S	90	AFBC	3.3	11	(84)	242	(0.56)	particulate	+	2,509	(5,532)	
0.9	32,100	(13,800)	M	75	AFBC	2.2	5.6	(47)	133	(0.31)	particulate	+	634	(1,400)	
0.9	32,100	(13,800)	I & S	83.9	AFBC	2.8	3.8	(30)	86	(0.20)	particulate	+	704	(1,550)	
0.9	32,100	(13,800)	S+	90	AFBC	3.3	2.3	(18)	52	(0.12)	particulate	+	758	(1,670)	
0.6	22,330	(9,600)	M	75	AFBC	2.2	5.7	(45)	128	(0.30)	particulate	+	667	(1,467)	
0.6	22,330	(9,600)	I & S	83.2	AFBC	2.7	3.8	(30)	86	(0.20)	particulate	+	722	(1,588)	
0.6	22,330	(9,600)	S+	90	AFBC	3.3	2.3	(18)	52	(0.12)	particulate	+	785	(1,727)	

TABLE 72d. AIR POLLUTION IMPACTS FROM "BEST" AND "COMMERCIALLY OFFERED" SO_2 CONTROL SYSTEMS FOR COAL-FIRED FBC BOILERS (58.6 MW_t or 200 × 10^6 Btu/hr heat input)

System							Air emissions						Secondary pollutants		
	Heat value				Type of control		SO_2				Other pollutants		Solid waste[#]		
%S	kJ/kg	(Btu/lb)	Control[†] level	Percent[‡] reduction	System	Ca/S[**]	g/s	(lb/h)	ng/J	(lb/10^6 Btu)	Pollutant	Effect[§]	"Best" system kg/h	(lb/h)	Commercial system kg/h (lb/h)
3.5	27,450	(11,800)	none	0	AFBC	0	142	(1,128)	2,245	(5.64)	NA	NA	NA		NA
0.9	32,100	(13,800)	none	0	AFBC	0	31	(248)	533	(1.24)	NA	NA	NA		NA
0.6	22,330	(9,600)	none	0	AFBC	0	30	(238)	512	(1.19)	NA	NA	NA		NA
3.5	27,450	(11,800)	M	78.7	AFBC	2.5	30	(240)	516	(1.20)	particulate	+	2,805	(6,181)	*
3.5	27,450	(11,800)	I	85	AFBC	2.9	21	(170)	364	(0.85)	particulate	+	3,080	(6,786)	*
3.5	27,450	(11,800)	S	90	AFBC	3.3	14	(112)	242	(0.56)	particulate	+	3,347	(7,376)	*
0.9	32,100	(13,800)	M	75	AFBC	2.2	7.8	(62)	133	(0.31)	particulate	+	846	(1,866)	*
0.9	32,100	(13,800)	I & S	83.9	AFBC	2.8	5.0	(40)	86	(0.20)	particulate	+	938	(2,066)	*
0.9	32,100	(13,800)	S+	90	AFBC	3.3	3.0	(24)	52	(0.12)	particulate	+	1,012	(2,227)	*
0.6	22,330	(9,600)	M	75	AFBC	2.2	7.6	(60)	128	(0.30)	particulate	+	887	(1,956)	*
0.6	22,330	(9,600)	I & S	83.2	AFBC	2.7	5.0	(40)	86	(0.20)	particulate	+	960	(2,117)	*
0.6	22,330	(9,600)	S+	90	AFBC	3.3	3.0	(24)	52	(0.12)	particulate	+	1,043	(2,303)	*

[*] These solid waste quantities are dependent upon the Ca/S molar feed ratio required for a given "commercially offered" system. Table in Section 3 gives a range of Ca/S ratios projected, and Table in Appendix gives the relative land use requirement for varying Ca/S ratios.

[†] M = moderate level
I = intermediate level
S = stringent level
S+ = greater than recommended stringent level

[‡] Variance from the 75 percent - moderate; 85 percent - intermediate; and 90 percent - stringent levels are due to the upper and lower limits of 516 ng/J (1.2 lb/MMBtu) and 86 ng/J (0.2 lb/MMBtu), respectively.

[§] + = an increase in emissions of the pollutant identified attributed to the SO_2 control method.
- = a decrease in emissions of the pollutant identified attributed to the SO_2 control method.

[#] These solid waste quantities were calculated as shown in Table of this section. The quantities of waste indicated in Table are based on the assumption that the sorbent fed is of average reactivity.

[**] Ca/S ratios based on a sorbent of average reactivity.

NA = Not Applicable.

The SO_2 emitted to the atmosphere is dependent upon the level of control which is exercised and the heat input rate of the boiler. For an 8.8 MW_t (30×10^3 Btu/hr) boiler using high sulfur coal (3.5 percent), the SO_2 emissions range from approximately 2.1 to 4.5 g/s (16 to 36 lb/hr) over the stringent to moderate control range. This compares to uncontrolled emissions of about 21 g/s (169 lb/hr) SO_2.

Table 72 also indicates that there is a slight increase in particulate emissions due to the control of SO_2 by limestone addition. To date, these results are not quantifiable; only trends in the data can be verified.

The effect of SO_2 control on NO_x emissions differs in that there is no predictable trend which can be identified. Depending on which operating variables are used to enhance SO_2 capture, NO_x emissions may increase or decrease. Generally, in a given system, once the design and operating conditions are established, increasing SO_2 capture will have little effect on NO_x emissions.

The largest potential impact of SO_2 control techniques in fluidized-bed combustion is the solid waste which is generated (spent bed material and carryover/fly ash) by the system. As SO_2 control levels are increased the amount of solid waste is increased. Table 72 shows the total quantity of solid waste generated for the three coals at three levels of control for the "best" system. Quantities of waste for the 8.8 MW_t (30×10^3 Btu/hr) boiler, assuming a sorbent of average reactivity, range from 128 kg/hr (281 lb/hr) for the lowest sulfur coal and SO_2 control level, to 502 kg/hr (1,105 lb/hr) for the highest sulfur coal and SO_2 control level.

As more research is done, SO_2 control via fluidized-bed combustion may be found to have beneficial effects beyond the SO_2 control itself. It is quite possible that as a better understanding of the chemical, physical and mechanical properties of the solid waste is developed, the material could have widespread use as a commercial byproduct (i.e., structural, road construction, agricultural, and soil conditioning materials). There are several research programs underway in this area whose initial results are very encouraging (see Section 6.2.2.5). If such uses of the solid waste find wide commercial application, a large degree of the adverse impact could translate into beneficial impact.

The environmental impact of AFBC solid waste is discussed in further detail in Section 6.2.2.

6.2.1.2 NO_x Emissions--

Table 73 illustrates NO_x emissions from coal-fired AFBC boilers under varying levels of NO_x control. NO_x emissions range from 1.9 to 2.6 g/s (15 to 21 lb/h) for the 8.8 MW_t (30×10^3 Btu/h) boiler, and 13 to 18 g/s (100 to 140 lb/h) for the 58.6 MW_t (200×10^3 Btu/h) boiler. It is assumed, based upon available data, that commercial-scale AFBC units will inherently be able to achieve all three levels of NO_x control, including the most stringent, without major adjustments to design and operating conditions.

6.2.1.3 Particulate Emissions--

Table 74 illustrates the air pollution impact of particulate control as applied to atmospheric fluidized-bed combustion. Uncontrolled particulate emissions from AFBC boilers range from 1.9 to 126 g/s (15 to 1,000 lb/h) in the boiler size range of 8.8 to 58.6 MW_t (30 to 200×10^3 Btu/hr). Moderate control levels of 107 ng/J (0.25 lb/10^6 Btu) and stringent control levels of 12.9 ng/J (0.03 lb/10^6 Btu), are expected to be achievable by AFBC with conventional add-on particulate control devices. The particulate material

TABLE 73. AIR POLLUTION IMPACTS FROM "BEST" NO_x CONTROL TECHNIQUES FOR COAL-FIRED, ATMOSPHERIC FLUIDIZED-BED COMBUSTION BOILERS

System					NO_x emissions				Other emissions		Secondary pollutants	
Heat rate		Fuel*	NO_x control level	Control method	g/s	(lb/h)	ng/J	(lb/10^6 Btu)	Pollutant	Degree of change	Beneficial	Adverse
MW	(10^6 Btu/h)											
8.8	(30)	Coal A, B & C	none	none	2.6	(21)	301	(0.7)	NA	NA	NA	NA
8.8	(30)	Coal A, B & C	M	AFBC	2.6	(21)	301	(0.7)	NA	NA	NA	NA
8.8	(30)	Coal A, B & C	I	AFBC	2.3	(18)	258	(0.6)	NA	NA	NA	NA
8.8	(30)	Coal A, B & C	S	AFBC	1.9	(15)	215	(0.5)	NA	NA	NA	NA
22	(75)	Coal A, B & C	none	none	6.7	(53)	301	(0.7)	NA	NA	NA	NA
22	(75)	Coal A, B & C	M	AFBC	6.7	(53)	301	(0.7)	NA	NA	NA	NA
22	(75)	Coal A, B & C	I	AFBC	5.7	(45)	254	(0.6)	NA	NA	NA	NA
22	(75)	Coal A, B & C	S	AFBC	4.8	(38)	251	(0.5)	NA	NA	NA	NA
44	(150)	Coal A, B & C	none	none	13	(105)	301	(0.7)	NA	NA	NA	NA
44	(150)	Coal A, B & C	M	AFBC	13	(105)	301	(0.7)	NA	NA	NA	NA
44	(150)	Coal A, B & C	I	AFBC	11	(90)	258	(0.6)	NA	NA	NA	NA
44	(150)	Coal A, B & C	S	AFBC	9.4	(75)	251	(0.5)	NA	NA	NA	NA
58.6	(200)	Coal A, B & C	none	none	18	(140)	301	(0.7)	NA	NA	NA	NA
58.6	(200)	Coal A, B & C	M	AFBC	18	(140)	301	(0.7)	NA	NA	NA	NA
58.6	(200)	Coal A, B & C	I	AFBC	15	(120)	258	(0.6)	NA	NA	NA	NA
58.6	(200)	Coal A, B & C	S	AFBC	13	(100)	251	(0.5)	NA	NA	NA	NA

*Coal A = High sulfur Eastern coal, 3.5 percent S; 10.6 percent Ash; 27,450 kJ/kg (11,800 Btu/lb).
Coal B = Low sulfur Eastern coal, 0.9 percent S; 6.9 percent Ash; 30,100 kJ/kg (13,800 Btu/lb).
Coal C = Subbituminous coal, 0.6 percent S; 5.4 percent Ash; 22,330 kJ/kg (9,600 Btu/lb).

NA = not applicable.

TABLE 74. AIR POLLUTION IMPACT FROM "BEST" PARTICULATE CONTROL TECHNIQUES FOR COAL-FIRED, ATMOSPHERIC FBC BOILERS

System						Particulate emissions				Other emissions		Secondary pollutants	
Heat Rate		Fuel*	Particulate control level	Percent particulate reduction	Type of particulate control device	g/s	(lb/h)	ng/J	(lb/10^6 Btu)	Pollutant	Degree of change	Beneficial	Adverse solid waste g/s (lb/h)
MW	(MMBtu/h)												
8.8	(30)	Coal A, B & C	none	0	none	1.9 - 18.9	(15 - 150)	215 - 2150	(0.5 - 5.0)	NA	NA	NA	NA
8.9	(30)	Coal A, B & C	M	50.0 - 95.3	MC, WS, ESP or FF	0.9	(7.5)	107	(0.25)	NA	NA	NA	0.9 - 18.0 (7.5 - 143)
8.8	(30)	Coal A, B & C	I	80.0 - 98.0	MC, WS, ESP or FF	0.4	(3.0)	43	(0.10)	NA	NA	NA	1.5 - 18.5 (12 - 147)
8.8	(30)	Coal A, B & C	S	93.3 - 99.3	ESP or FF	0.1	(0.9)	12.9	(0.03)	NA	NA	NA	1.8 - 18.8 (14 - 149)
22	(75)	Coal A, B & C	none	0	none	4.7 - 47.2	(38 - 375)	215 - 2150	(0.5 - 5.0)	NA	NA	NA	NA
22	(75)	Coal A, B & C	M	50.0 - 94.9	MC, WS, ESP or FF	2.4	(19)	107	(0.25)	NA	NA	NA	2.4 - 45 (19 - 356)
22	(75)	Coal A, B & C	I	81.6 - 98.1	MC, WS, ESP or FF	0.9	(7.5)	43	(0.10)	NA	NA	NA	3.9 - 46 (31 - 368)
22	(75)	Coal A, B & C	S	94.7 - 99.5	ESP or FF	0.3	(2.3)	12.9	(0.03)	NA	NA	NA	4.5 - 47 (36 - 373)
44	(150)	Coal A, B & C	none	0	none	9.4 - 94.5	(75 - 750)	215 - 2150	(0.5 - 5.0)	NA	NA	NA	NA
44	(150)	Coal A, B & C	M	50.0 - 94.9	MC, WS, ESP or FF	4.8	(38)	107	(0.25)	NA	NA	NA	4.7 - 90 (37 - 712)
44	(150)	Coal A, B & C	I	80.0 - 98.0	MC, WS, ESP or FF	1.9	(15)	43	(0.10)	NA	NA	NA	7.6 - 93 (60 - 735)
44	(150)	Coal A, B & C	S	94.7 - 99.5	ESP or FF	5.7	(4.5)	12.9	(0.03)	NA	NA	NA	8.9 - 94 (71 - 746)
58.6	(200)	Coal A, B & C	none	0	none	12.6 - 126.0	(100 - 1000)	215 - 2150	(0.5 - 4.0)	NA	NA	NA	NA
58.6	(200)	Coal A, B & C	M	50.0 - 95.0	MC, WS, ESP or FF	6.3	(50)	107	(0.25)	NA	NA	NA	6.3 - 120 (50 - 950)
58.6	(200)	Coal A, B & C	I	80.0 - 98.0	MC, WS, ESP or FF	2.5	(20)	43	(0.10)	NA	NA	NA	10 - 123 (80 - 980)
58.6	(200)	Coal A, B & C	S	94.0 - 99.4	ESP or FF	0.8	(6.0)	12.9	(0.03)	NA	NA	NA	12 - 125 (94 - 994)

[*] Coal A = High sulfur Eastern coal, 27,450 kJ/kg (11,800 Btu/lb); 3.5 percent S; 10.6 percent A
Coal B = Low sulfur Eastern coal, 32,100 kJ/kg (13,800 Btu/lb); 0.9 percent S; 6.9 percent A
Coal C = Subbituminous coal, 22,230 kJ/kg (9,600 Btu/lb); 0.6 percent S; 5.4 percent A

[†] FF - Fabric filter
ESP - Electrostatic precipitator
WS - Wet scrubber
MC - Multitube cyclone

[‡] These levels of particulate emissions are based on the following proposed standards:

Moderate - 107 ng/J (0.25 lb/10^6 Btu)
Intermediate - 43 ng/J (0.10 lb/10^6 Btu)
Stringent - 12.9 ng/J (0.03 lb/10^6 Btu)

The ranges of the amount of particulate emissions are based on the range of material elutriated from the bed which is dependent upon Ca/S ratio, gas residence time and velocity, particle size and particle size distribution.

which is collected ranges from 0.9 to 125 g/s (7.5 to 994 lb/hr) depending upon the boiler size and level of SO_2 and particulate control. The 125 g/s (994 lb/hr) of collected particulates compares to 113 g/s (900 lb/hr) from a conventional system of equivalent coal usage. A larger quantity of particulates is expected from FBC than from conventional systems as a result of the attrited bed material in the carryover. The particulates collected comprise from 5 to 15 percent of the total solid waste from AFBC. The environmental impact of the combined solid waste is covered separately in Section 6.2.2 of this report.

6.2.1.4 Trace Element Emissions--

The emissions of trace elements from coal-fired fluidized-bed combustion systems on an industrial scale have not been documented. To date, there is no reason to suspect that trace element emissions from fluidized-bed combustion should be worse than that encountered in any coal-fired system. In fact, the lower temperatures of FBC combustion may reduce the tendency of some of the more volatile elements to be enriched on the finer fly ash particulates, a phenomenon which is sometimes encountered in conventional coal-fired systems. In bench scale experiments on a 6-in. pressurized combustor, Argonne reported trace element emissions which were lower than what one would expect from conventional systems.[6] A preliminary environmental assessment by GCA Corporation concluded that coal-fired FBC should present no problems for airborne trace element emissions.[7] However, it is important to note that any conclusions to date on FBC trace element emissions are based on limited laboratory scale data. Further experimental verification of the characteristics of trace metals in air emissions (and solid waste) is necessary on industrial-scale FBC systems.

6.2.2 Solid Waste

The major adverse environmental impact of fluidized-bed combustion is expected to be the solid waste which it produces. Solid residue from the fluidized-bed process consists of spent bed material (largely calcined and sulfated sorbent), and a mixture of fly ash collected in the particulate control devices.

6.2.2.1 Quantity of Solid Waste Generated--

The amount of solid waste material produced is a function of the fuel and sorbent characteristics. The following major variables are considered in estimating the amount of solid waste which will be generated.

- Ca/S molar feed ratio required
    - reactivity of the sorbent type (categorized by chemical and physical properties)
    - design/operating conditions which affect sorbent performance (sorbent particle size and gas phase residence time, etc.)
    - percent SO_2 reduction required
- fuel sulfur
- fuel ash
- fuel heating value

Different sorbents have varying calcium contents and calcium utilization rates. Once a sorbent is chosen for a specific application, the calcium utilization rate can generally be increased by reducing the particle size. The design gas velocity and bed height can then be adjusted to give the optimum gas-solids contact time for a given particle size. The greater the gas phase residence time is, the greater the calcium utilization. Once these parameters are established, the fuel feed and the level of control to be met determine the sorbent mass feed rate and amount of solid waste generated in the fluidized bed.

To indicate the environmental impact of the waste, Tables 75a through 75d demonstrate the waste produced at varying control levels using different coals. The methods used to calculate the mass and composition of the total waste are indicated in the footnotes under Table 75d. For each boiler size, coal type and control level, a range of waste production rates is given, representing the expected range of sorbent reactivities. Solid waste for an 8.8 MW_t (30 × 10^3 Btu/hr) thermal input coal-fired boiler ranges from 115 to 580 kg/h (255 to 1,278 lb/h). At 58.6 MW_t (200 × 10^3 Btu/hr), the estimated maximum waste is 3,873 kg/h (8,533 lb/h). These solid waste loadings constitute the total waste produced by the system; about 85 to 95 percent of the waste will be withdrawn as spent bed material, assuming that the material collected in the primary cyclone is recycled to the bed. The remaining 5 to 15 percent elutriates from the bed, passes through the primary cyclone, and is collected by a final particulate control device.

At levels of control specified earlier in Table 74 for particulate emissions, nearly all the particulate matter is collected, and it is assumed to be mixed with the spent bed material for disposal.

The land use for disposal of the solid waste has been projected using the sensitivity analysis program discussed in Appendix C. Table C-29 in the Appendix shows the variation of disposal area needed for wastes from AFBC and conventional-fired boilers with NO_x and SO_2 control. The table shows the impact for the four boiler sizes, using the three coals which have been represented throughout the report, and the optional SO_2 control levels. Figure 55 illustrates the effect of these variables upon the land requirements for an FBC site where high sulfur (3.5 percent S) coal is burned. For each boiler capacity the

TABLE 75a. SOLID WASTE GENERATED BY A ONCE-THROUGH, LIMESTONE-FED, COAL-FIRED, "BEST SYSTEM" ATMOSPHERIC FBC BOILER (8.8 MW or 30 × 10^6 Btu/hr heat input)

% Sulfur	% Ash	% SO$_2$ control	Level of* control	Ca/S† ratio	Limestone feed‡ kg/h (lb/h)	Inerts from limestone (10%) kg/h (lb/h)	Uncalcined CaCO$_3$§ kg/h (lb/h)	Unreacted CaO‖ kg/h (lb/h)	CaSO$_4$ generated** kg/h (lb/h)	Coal ash†† kg/h (lb/h)	Total solid waste kg/h (lb/h)
3.5	10.6	78.7	M	2.5	350 (772)	35 (77)	16 (35)	113 (248)	134 (296)	122 (269)	420 (925)
				3.4	478 (1,050)	48 (105)	22 (47)	174 (381)	134 (296)	122 (269)	500 (1,098)
				1.8	252 (556)	25 (56)	11 (25)	66 (144)	134 (296)	122 (269)	358 (790)
3.5	10.6	85	I	2.9	406 (895)	41 (90)	18 (40)	134 (297)	146 (320)	122 (289)	461 (1,016)
				3.8	533 (1,173)	53 (117)	24 (53)	195 (430)	146 (320)	122 (269)	540 (1,189)
				2.1	294 (648)	29 (65)	13 (29)	81 (178)	146 (320)	122 (269)	391 (861)
3.5	10.6	90	S	3.3	463 (1,019)	46 (102)	21 (46)	158 (348)	155 (340)	122 (269)	502 (1,105)
				4.2	588 (1,296)	59 (130)	26 (58)	218 (481)	155 (340)	122 (269)	580 (1,278)
				2.3	322 (710)	32 (71)	14 (32)	90 (200)	155 (340)	122 (269)	413 (912)
0.9	6.9	75	M	2.2	68 (150)	68 (15)	3.1 (6.8)	21 (46)	29 (63)	68 (150)	128 (281)
				3.2	99 (218)	9.9 (22)	4.5 (9.8)	35 (78)	29 (63)	68 (150)	146 (323)
				1.6	49 (109)	4.9 (11)	2.2 (4.9)	11 (26)	29 (63)	68 (150)	115 (255)
0.9	6.9	83.9	I & S	2.8	87 (191)	8.7 (19)	3.9 (8.6)	29 (62)	32 (70)	68 (150)	142 (310)
				3.7	114 (252)	1.1 (25)	5.0 (11)	42 (92)	32 (70)	68 (150)	158 (348)
				2.0	62 (136)	6.2 (14)	2.8 (6.1)	17 (36)	32 (70)	68 (150)	126 (276)
0.9	6.9	90	S +	3.3	102 (225)	10 (23)	4.6 (10)	35 (77)	34 (75)	68 (150)	152 (335)
				4.2	130 (286)	13 (29)	5.9 (13)	48 (106)	34 (75)	68 (150)	169 (373)
				2.3	71 (157)	7.1 (16)	3.2 (7.1)	20 (44)	34 (75)	68 (150)	132 (292)
0.6	5.4	75	M	2.2	65 (143)	6.5 (14)	2.9 (6.4)	20 (43)	27 (61)	77 (169)	133 (293)
				3.2	94 (208)	9.4 (21)	4.2 (9.4)	34 (75)	27 (61)	77 (169)	152 (335)
				1.6	47 (104)	4.7 (10)	2.1 (4.7)	12 (25)	27 (61)	77 (169)	123 (270)
0.6	5.4	83.2	I & S	2.7	80 (176)	8.0 (18)	3.6 (7.9)	26 (57)	29 (66)	77 (169)	144 (318)
				3.6	106 (234)	11 (23)	4.8 (11)	39 (85)	29 (66)	77 (169)	161 (354)
				2.0	59 (130)	5.9 (13)	2.7 (5.9)	16 (35)	29 (66)	77 (169)	131 (289)
0.6	5.4	90	S +	3.3	98 (215)	9.8 (22)	4.4 (9.7)	34 (74)	32 (70)	77 (169)	157 (345)
				4.2	124 (273)	12 (27)	5.6 (12)	46 (102)	32 (70)	77 (169)	173 (380)
				2.3	68 (150)	6.8 (15)	3.1 (6.8)	20 (43)	32 (70)	77 (169)	139 (304)

TABLE 75b. SOLID WASTE GENERATED BY A ONCE-THROUGH, LIMESTONE-FED, COAL-FIRED, "BEST SYSTEM" ATMOSPHERIC FBC BOILER (22 MW or 75 × 10^6 Btu/hr heat input)

% Sulfur	% Ash	% SO_2 control	Level of* control	Ca/S† ratio	Limestone feed‡ kg/h (lb/h)	Inerts from limestone (10%) kg/h (lb/h)	Uncalcined $CaCO_3$§ kg/h (lb/h)	Unreacted CaO# kg/h (lb/h)	$CaSO_4$ generated** kg/h (lb/h)	Coal ash†† kg/h (lb/h)	Total solid waste kg/h (lb/h)
3.5	10.6	78.7	M	2.5	877 (1,932)	88 (193)	39 (87)	281 (618)	338 (746)	306 (674)	1,052 (2,318)
				3.4	1,193 (2,628)	119 (263)	54 (118)	432 (951)	338 (746)	306 (674)	1,249 (2,752)
				1.8	632 (1,391)	63 (139)	28 (62)	163 (359)	338 (746)	306 (674)	898 (1,980)
3.5	10.6	85	I	2.9	1,017 (2,241)	102 (224)	46 (101)	337 (742)	364 (804)	306 (674)	1,155 (2,545)
				3.8	1,333 (2,937)	133 (294)	60 (132)	488 (1,075)	364 (804)	306 (674)	1,351 (2,979)
				2.1	737 (1,623)	74 (162)	33 (73)	203 (446)	364 (804)	306 (674)	980 (2,159)
3.5	10.6	90	S	3.3	1,158 (2,551)	116 (255)	52 (115)	395 (870)	386 (852)	306 (674)	1,255 (2,766)
				4.2	1,474 (3,246)	147 (325)	66 (146)	547 (1,203)	386 (852)	306 (674)	1,452 (3,200)
				2.3	807 (1,178)	81 (178)	36 (80)	227 (500)	386 (852)	306 (674)	1,036 (2,284)
0.9	6.9	75	M	2.2	170 (374)	17 (37)	4.8 (17)	52 (115)	70 (155)	170 (375)	314 (699)
				3.2	247 (544)	25 (54)	11 (24)	89 (196)	70 (155)	170 (375)	365 (804)
				1.6	123 (272)	12 (27)	5.5 (12)	30 (66)	70 (155)	170 (375)	288 (635)
0.9	6.9	83.9	I & S	2.8	216 (476)	22 (48)	9.7 (21)	71 (156)	80 (175)	170 (375)	352 (775)
				3.7	285 (629)	29 (63)	13 (28)	104 (229)	80 (175)	170 (375)	396 (870)
				2.0	154 (340)	15 (34)	6.9 (15)	41 (91)	80 (175)	170 (375)	313 (690)
0.9	6.9	90	S +	3.3	255 (561)	26 (56)	11 (25)	87 (192)	85 (187)	170 (375)	379 (835)
				4.2	324 (714)	32 (71)	15 (32)	120 (265)	85 (187)	170 (375)	422 (930)
				2.3	178 (391)	18 (39)	8.0 (18)	50 (110)	85 (187)	170 (375)	331 (729)
0.6	5.4	75	M	2.2	163 (358)	16 (36)	7.3 (16)	50 (109)	68 (151)	191 (422)	332 (734)
				3.2	237 (521)	24 (52)	11 (23)	85 (187)	68 (151)	191 (422)	379 (835)
				1.6	118 (261)	12 (26)	5.3 (12)	29 (63)	68 (151)	191 (422)	305 (674)
0.6	5.4	83.2	I & S	2.7	200 (440)	20 (44)	9.0 (20)	65 (143)	75 (165)	191 (422)	360 (794)
				3.6	266 (587)	27 (59)	12 (26)	97 (213)	75 (165)	191 (422)	402 (885)
				2.0	148 (326)	15 (33)	6.7 (15)	40 (88)	75 (165)	191 (422)	328 (723)
0.6	5.4	90	S +	3.3	244 (538)	24 (54)	11 (24)	83 (184)	83 (180)	191 (422)	392 (864)
				4.2	311 (684)	31 (68)	14 (31)	114 (253)	83 (180)	191 (422)	433 (954)
				2.3	170 (375)	17 (38)	7.7 (17)	48 (106)	83 (180)	191 (422)	347 (763)

TABLE 75c. SOLID WASTE GENERATED BY A ONCE-THROUGH, LIMESTONE-FED, COAL-FIRED, "BEST SYSTEM" ATMOSPHERIC FBC BOILER (44 MW or 150 × 10^6 Btu/hr heat input)

% Sulfur	% Ash	% SO_2 control	Level of* control	Ca/S† ratio	Limestone feed‡ kg/h (lb/h)	Inerts from limestone (10%) kg/h (lb/h)	Uncalcined $CaCO_3$§ kg/h (lb/h)	Unreacted CaO¶ kg/h (lb/h)	$CaSO_4$ generated** kg/h (lb/h)	Coal ash†† kg/h (lb/h)	Total solid waste kg/h (lb/h)
3.5	10.6	78.7	M	2.5	1,754 (3,865)	175 (387)	79 (174)	562 (1,236)	675 (1,491)	611 (1,347)	2,102 (4,635)
				3.4	2,386 (5,255)	239 (526)	107 (236)	864 (1,902)	675 (1,491)	611 (1,347)	2,496 (5,502)
				1.8	1,264 (2,782)	126 (278)	56 (125)	326 (718)	675 (1,491)	611 (1,347)	1,794 (3,959)
3.5	10.6	85	I	2.9	2,034 (4,482)	203 (448)	92 (202)	674 (1,484)	729 (1,608)	611 (1,347)	2,309 (5,089)
				3.8	2,666 (5,874)	267 (587)	120 (264)	976 (2,150)	729 (1,608)	611 (1,347)	2,703 (5,956)
				2.1	1,474 (3,246)	147 (325)	66 (146)	406 (892)	729 (1,608)	611 (1,347)	1,959 (4,318)
3.5	10.6	90	S	3.3	2,316 (5,102)	232 (510)	104 (230)	790 (1,740)	772 (1,705)	611 (1,347)	2,509 (5,532)
				4.2	2,948 (6,492)	295 (649)	133 (292)	1,094 (2,406)	772 (1,705)	611 (1,347)	2,905 (6,399)
				2.3	1,614 (3,556)	161 (356)	73 (160)	454 (1,000)	772 (1,705)	611 (1,347)	2,071 (4,568)
0.9	6.9	75	M	2.2	340 (748)	34 (75)	15 (34)	104 (230)	141 (311)	340 (750)	634 (1,400)
				3.2	494 (1,086)	49 (109)	22 (49)	178 (392)	141 (311)	340 (750)	730 (1,611)
				1.6	246 (544)	25 (54)	11 (24)	60 (132)	141 (311)	340 (750)	577 (1,271)
0.9	6.9	83.9	I & S	2.8	432 (952)	43 (95)	19 (43)	142 (312)	160 (350)	340 (750)	704 (1,550)
				3.7	570 (1,258)	57 (126)	26 (57)	208 (458)	160 (350)	340 (750)	791 (1,741)
				2.0	308 (680)	31 (68)	14 (31)	82 (182)	160 (350)	340 (750)	627 (1,381)
0.9	6.9	90	S +	3.3	510 (1,122)	51 (112)	23 (50)	174 (384)	170 (374)	340 (750)	758 (1,670)
				4.2	648 (1,428)	65 (143)	29 (64)	240 (530)	170 (374)	340 (750)	844 (1,861)
				2.3	356 (782)	36 (78)	16 (35)	100 (220)	170 (374)	340 (750)	662 (1,457)
0.6	5.4	75	M	2.2	326 (716)	33 (72)	15 (32)	100 (218)	136 (301)	383 (844)	667 (1,467)
				3.2	474 (1,042)	47 (104)	21 (47)	170 (374)	136 (301)	383 (844)	757 (1,670)
				1.6	236 (522)	24 (52)	11 (23)	58 (126)	136 (301)	383 (844)	612 (1,346)
0.6	5.4	83.2	I & S	2.7	400 (880)	40 (88)	18 (40)	130 (286)	151 (330)	383 (844)	722 (1,588)
				3.6	532 (1,174)	53 (117)	24 (53)	194 (426)	151 (330)	383 (844)	805 (1,770)
				2.0	296 (652)	30 (65)	13 (29)	80 (176)	151 (330)	383 (844)	657 (1,444)
0.6	5.4	90	S +	3.3	488 (1,076)	49 (108)	22 (48)	166 (368)	165 (359)	383 (844)	785 (1,727)
				4.2	622 (1,368)	62 (137)	28 (62)	228 (506)	165 (359)	383 (844)	866 (1,908)
				2.3	340 (750)	34 (75)	15 (34)	96 (212)	165 (359)	383 (844)	693 (1,524)

TABLE 75d. SOLID WASTE GENERATED BY A ONCE-THROUGH, LIMESTONE-FED, COAL-FIRED, "BEST SYSTEM" ATMOSPHERIC FBC BOILER (58.6 MW or 200 × 10^6 Btu/hr heat input)

% Sulfur	% Ash	% SO_2 control	Level of* control	Ca/S† ratio	Limestone feed† kg/h (lb/h)	Inerts from limestone (10%) kg/h (lb/h)	Uncalcined $CaCO_3$§ kg/h (lb/h)	Unreacted CaO# kg/h (lb/h)	$CaSO_4$ generated** kg/h (lb/h)	Coal ash†† kg/h (lb/h)	Total solid waste kg/h (lb/h)
3.5	10.6	78.7	M	2.5	2,339 (5,153)	234 (515)	105 (232)	748 (1,648)	903 (1,989)	815 (1,797)	2,805 (6,181)
				3.4	3,181 (7,007)	318 (701)	143 (315)	1,151 (2,536)	903 (1,989)	815 (1,797)	3,330 (7,338)
				1.8	1,684 (3,709)	168 (371)	76 (167)	434 (957)	903 (1,989)	815 (1,797)	2,396 (5,281)
3.5	10.6	85	I	2.9	2,713 (5,976)	271 (598)	122 (269)	898 (1,978)	974 (2,144)	815 (1,797)	3,080 (6,786)
				3.8	3,555 (7,832)	356 (783)	160 (352)	1,301 (2,866)	974 (2,144)	815 (1,797)	3,606 (7,942)
				2.1	1,965 (4,328)	196 (433)	88 (195)	540 (1,189)	974 (2,144)	815 (1,797)	2,613 (5,758)
3.5	10.6	90	S	3.3	3,089 (6,803)	308 (680)	139 (306)	1,053 (2,320)	1,032 (2,273)	815 (1,797)	3,347 (7,376)
				4.2	3,930 (8,656)	393 (865)	177 (390)	1,456 (3,208)	1,032 (2,273)	915 (1,797)	3,873 (8,533)
				2.3	2,152 (4,741)	215 (474)	97 (213)	605 (1,333)	1,032 (2,273)	815 (1,797)	2,764 (6,090)
0.9	6.9	75	M	2.2	453 (997)	45 (100)	20 (45)	139 (306)	188 (415)	454 (1,000)	846 (1,866)
				3.2	657 (1,448)	66 (145)	29 (65)	237 (522)	188 (415)	454 (1,000)	974 (2,147)
				1.6	329 (725)	33 (73)	15 (33)	80 (176)	188 (415)	454 (1,000)	770 (1,697)
0.9	6.9	83.9	I & S	2.8	576 (1,269)	58 (127)	26 (57)	189 (416)	211 (466)	454 (1,000)	938 (2,066)
				3.7	761 (1,677)	76 (168)	34 (75)	278 (611)	211 (466)	454 (1,000)	1,053 (2,320)
				2.0	412 (907)	41 (91)	19 (41)	110 (243)	211 (466)	454 (1,000)	835 (1,841)
0.9	6.9	90	S +	3.3	679 (1,496)	68 (150)	31 (67)	233 (512)	226 (498)	454 (1,000)	1,012 (2,227)
				4.2	864 (1,904)	86 (190)	39 (86)	321 (707)	226 (498)	454 (1,000)	1,126 (2,481)
				2.3	474 (1,043)	47 (104)	21 (47)	134 (294)	226 (498)	454 (1,000)	882 (1,943)
0.6	5.4	75	M	2.2	434 (955)	43 (96)	20 (43)	132 (291)	182 (401)	510 (1,125)	887 (1,956)
				3.2	631 (1,389)	63 (139)	28 (63)	226 (499)	182 (401)	510 (1,125)	1,009 (2,227)
				1.6	316 (696)	32 (70)	14 (31)	76 (168)	182 (401)	510 (1,125)	814 (1,795)
0.6	5.4	83.2	I & S	2.7	533 (1,173)	53 (117)	24 (53)	174 (382)	199 (440)	510 (1,125)	960 (2,117)
				3.6	711 (1,565)	71 (157)	32 (70)	258 (568)	199 (440)	510 (1,125)	1,070 (2,360)
				2.0	395 (869)	40 (87)	18 (39)	107 (235)	199 (440)	510 (1,125)	874 (1,926)
0.6	5.4	90	S +	3.3	651 (1,435)	65 (144)	29 (65)	223 (491)	216 (478)	510 (1,125)	1,043 (2,303)
				4.2	828 (1,824)	83 (182)	37 (82)	307 (675)	216 (478)	510 (1,125)	1,153 (2,542)
				2.3	454 (1,000)	45 (100)	20 (45)	129 (283)	216 (478)	510 (1,125)	920 (2,031)

TABLE 75 (continued)

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[*] M = moderate level
I = intermediate level
S = stringent level
S+ = greater than recommended stringent level

[†] Each level of control is shown to have three Ca/S ratios assocated with it. This range of ratios represents the projected range of sorbent feed rates (for the "best system" design for SO_2 control) resulting from the expected range of sorbent reactivities.

[‡] Limestone - assumed 90 percent $CaCO_3$; 10 percent inerts.

[§] 95 percent of the $CaCO_3$ is assumed to be calcined.

[#] Unreacted CaO = CaO produced - CaO used;

$$\text{Rate CaO produced} = \text{percent } CaCO_3 \text{ in feed} \times \text{percent } CaCO_3 \text{ calcined} \times \frac{\text{molecular weight of CaO}}{\text{molecular weight of } CaCO_3} \times \text{limestone feed rate;}$$

$$\left(\text{i.e., } 0.90 \times 0.95 \times \frac{56}{100} \times \text{limestone feed}\right)$$

$$\text{Rate CaO used} = \text{fractional } SO_2 \text{ control level} \times \text{rate } SO_2 \text{ released by coal combustion (kg/h or lb/h)} \times \frac{\text{molecular weight of CaO}}{\text{molecular weight of } SO_2} \quad \left(\text{i.e. } \frac{56}{64}\right)$$

[**] $$CaSO_4 \text{ generated} \times \text{CaO used} \times \frac{\text{molecular weight of CaSO}}{\text{molecular weight of CaO}} \quad \left(\text{i.e. } \frac{136}{56}\right)$$

[††] Total solid waste quantities include 100 percent of the coal ash, regardless of whether the ash is withdrawn from the bed or captured in primary and final fly ash control devices. Similarily, the total waste includes all of the spent sorbent regardless whether the spent sorbent is withdrawn from the bed.

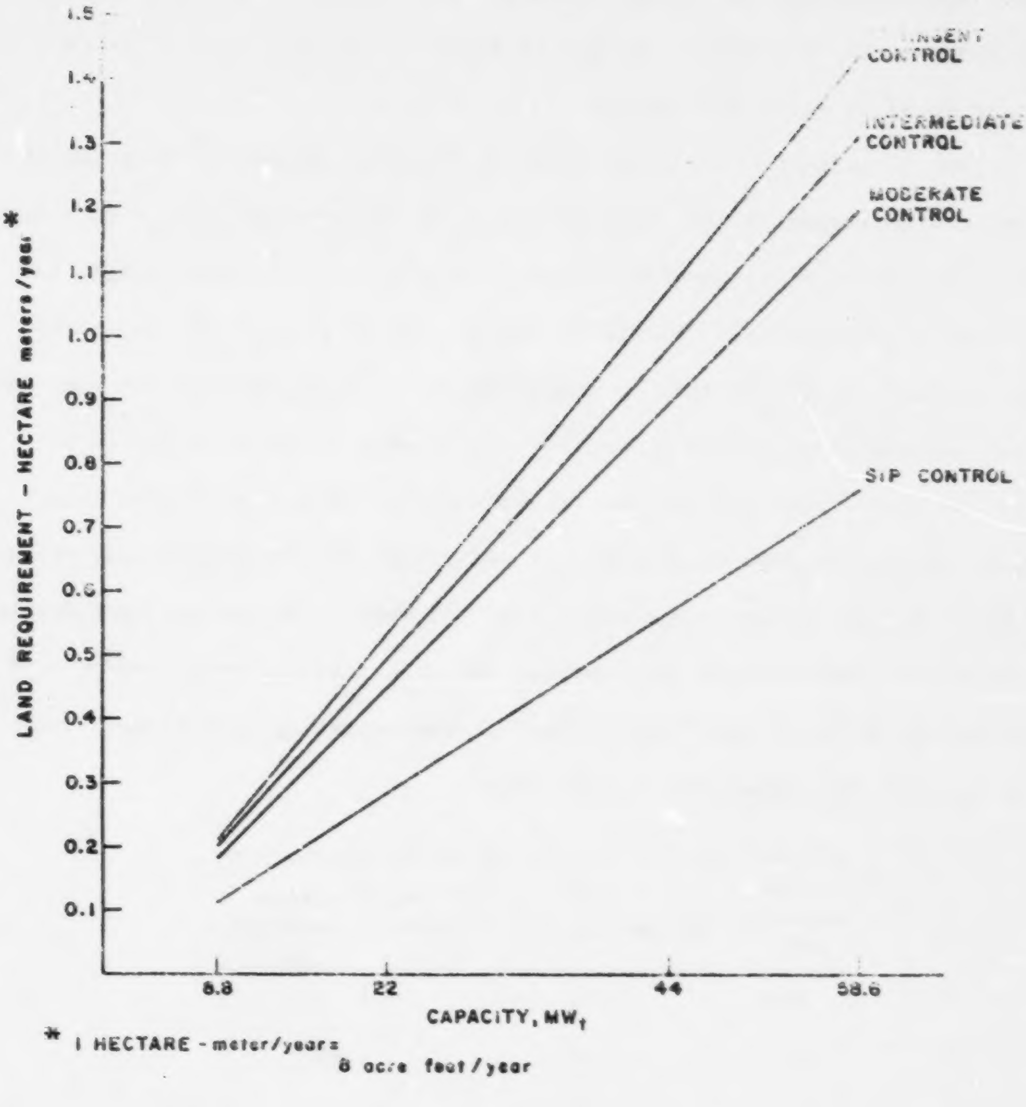
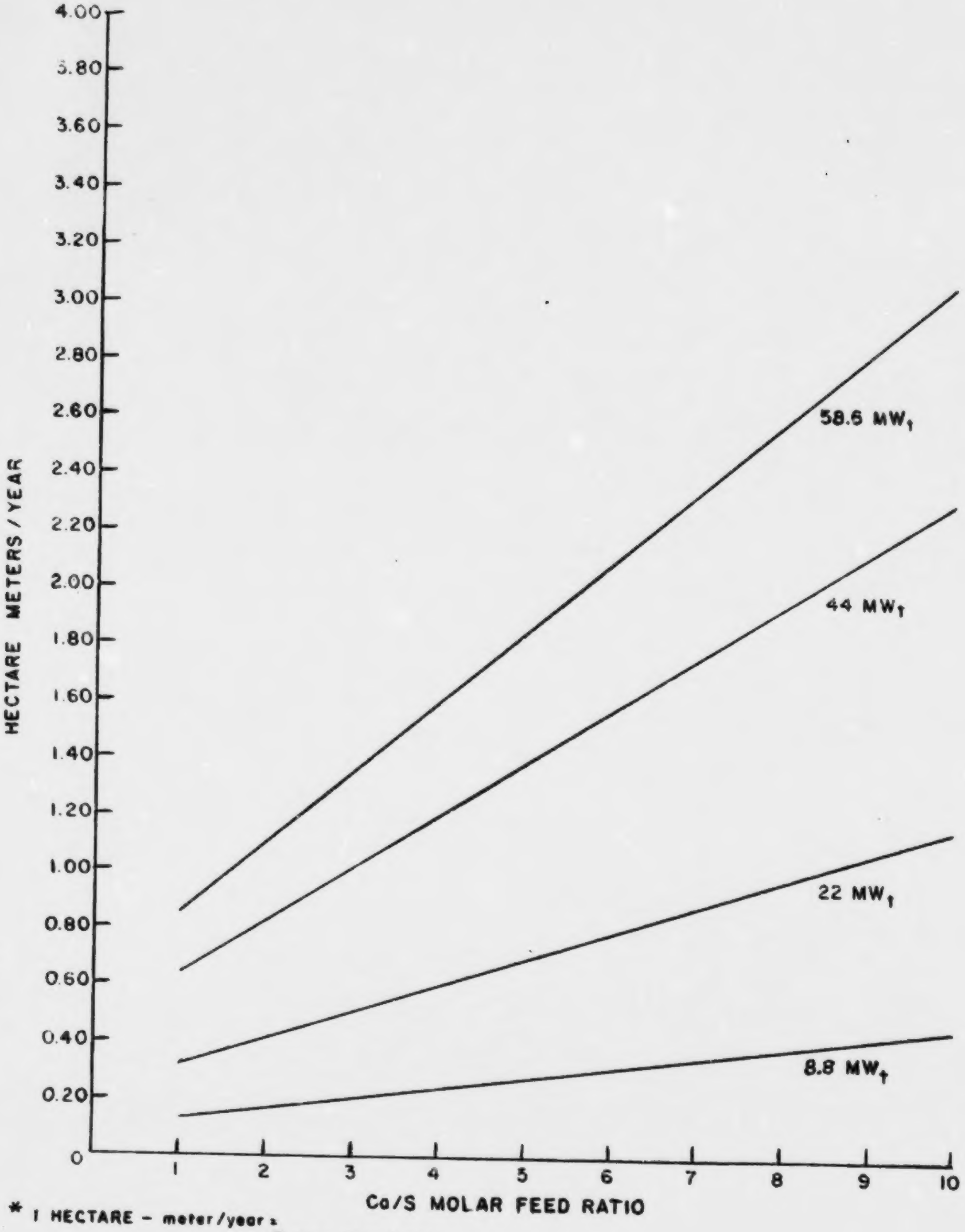


Figure 55. Land requirements for FBC burning high sulfur coal using medium reactivity limestone.

land use requirements of the stringent level of SO_2 control are nearly twice the requirements of that which would result if SO_2 control were equal to current SIP requirements. This impact is slight for the small boilers, but can be significant for the larger boilers. The range of land needed for solid waste disposal for the boiler systems in Figure 55 is 0.11 to 1.43 hectare meter/yr (0.92 to 11.56 acre ft/yr).

Figure 56 indicates the sensitivity of the land needed for disposal of the waste with respect to the Ca/S ratio, as it is increased from 1 to 9.9. (Ca/S ratios above 6 are considered unrealistic but are provided simply for the reader's perspective.) Figure 56 can be used to project the impact of "commercially offered" systems by comparing the Ca/S ratios used for the commercial systems as discussed in Section 3.0 of this report with the Ca/S ratios and associated land use requirements projected in the figure. For example, taking the case of the Babcock and Wilcox (U.S.) commercially offered design burning an Eastern high sulfur coal and Western 90 percent Ca/Limestone with stringent (90 percent) SO_2 control, the Ca/S ratio is projected at 4.58, as opposed to 2.83 for the "best system" design/operating conditions. The following land use comparisons can be made:

Boiler capacity MW_t	B&W hectare meter/yr	"best" system hectare meter/yr
8.8	0.26	0.20
22	0.66	0.49
44	1.31	0.78
58.6	1.74	1.30



* 1 HECTARE - meter/year = 8 acre feet/year

Figure 56. Land use requirements for disposal of solid waste.

6.2.2.2 Environmental Properties of FBC Solid Waste--

Disposal of solid waste from FBC systems is expected to occur by landfilling the material. The environmental impact of this method of disposal is under investigation. The primary sources of environmental degradation are the leachate formed by rainwater runoff and percolation after landfilling, and the heat release from the material upon initial contact with water, due to hydration of the CaO in the waste.

The disposal of solid waste is governed by laws promulgated under the auspices of the 1976 Resource Conservation and Recovery Act (RCRA, PL 94-580). In response to RCRA, the EPA has proposed a regulatory program to manage and control the nation's hazardous wastes from generation to disposal. The program includes criteria for identification of hazardous wastes (toxic, corrosive, ignitable and reactive), and rules and regulations for their management and control. When the EPA proposed the hazardous waste regulations, it set aside a unique category of special wastes - certain large volume wastes of which portions would be hazardous. The EPA plans to propose regulations governing special wastes in the early part of 1982. Until that time, the EPA has prepared special standards for each type of special waste. Although solid residues from coal-fired fluidized-bed combustion systems have not been regulated by the proposed program, they are a potential candidate for inclusion in the special wastes category which will include cement kiln dust, utility wastes (fly ash, bottom ash, scrubber sludge), phosphate mining and processing wastes, uranium mining wastes, other mining wastes, and oil and gas drilling muds and oil production brines.

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FBC residue seems to be a potential candidate for the special waste category because it will be generated in large quantities once the FBC technology is commercialized. Also, it contains similar chemical constituents to those found in utility wastes and cement kiln dust.

The EPA recommends that only the hazardous portions of special wastes comply with the proposed special standards. Section 3001 of the regulation provides the means for determining whether a waste is hazardous for the purpose of the Act. The hazardous portions of solid residues of FBC systems will be determined by testing by toxicity - one of the characteristics exhibited by hazardous wastes when improperly disposed of. A waste is considered toxic for the purpose of the Act if a chemical analysis of its water extract obtained in accordance with the Extraction Procedure (EP) reveals the presence of one of certain chemicals in concentrations which exceed ten times the drinking water standards. The contaminants and their maximum allowable concentrations are listed below:

Contaminant	Maximum allowable extract level (mg/l)
Arsenic	0.50
Barium	10.0
Cadmium	0.10
Chromium	0.50
Lead	0.50
Mercury	0.02
Selenium	0.10
Silver	0.50
Endrin	0.002
Lindane	0.040
Methoxychlor	1.0
Toxaphene	0.050
2,4-P	1.0
2,4,5-TP	0.10

This list is not final and may be revised by EPA through the rulemaking process as information develops. Since the last six contaminants are synthetic organic compounds, it is very likely that they will be present in leachates from FBC wastes. The analysis, therefore, can be limited to the metal ions portions of the list. Wastes from several small scale fluidized-bed combustion units have recently been tested by Westinghouse for EPA. None of the eight metals listed above exceeded the maximum allowable extract level.[8]

If these pollutants are measured in concentrations above the maximum allowable extract level, then the solid waste must be disposed of in compliance with the rules and regulations set forth in RCRA for toxic wastes. FBC residue will probably not be considered a candidate for corrosive, ignitable or reactive categories under RCRA.

Current interpretations of RCRA indicate that "corrosive" applies to liquid wastes and not leachate from solid waste; hence, despite its high pH, FBC waste would probably not be considered "corrosive." Furthermore, even though it does release heat upon exposure to water, the reaction does not seem sufficient to meet current EPA criteria for "reactive" waste.

The British Coal Utilization Research Administration (BCURA),[9] Pope, Evans, and Robbins,[10] Westinghouse,[11,12] and Ralph Stone and Company,[13] have conducted laboratory tests to investigate the properties of the leachate obtained from the coal ash/limestone waste using distilled water. Their test results generally showed the following common factors:

- high calcium content;
- high sulfate content;
- high total dissolved solids, due to $CaSO_4$ going into solution; and,
- high pH (10 to 12) due to CaO content.

One of the most definitive evaluations of the potential contamination from FBC waste was done by Westinghouse Research Laboratories.[14,15] Leachates were generated using distilled, deionized water in laboratory shake tests for a variety of FBC wastes;* the resulting leachate concentrations were then compared with drinking water standards (National Interim Primary Drinking Water Regulations (NIPDWR-1975), U.S. Public Health Service (USPHS) Drinking Standards). The data are summarized in Table 76. This is a very conservative approach and would tend to overestimate the impact since: (1) the laboratory shake tests are designed to maximize the water extraction forces; and (2) direct comparison with drinking water standards does not allow for any dilution of a leachate plume in the ground water. It is also important to note that drinking water standards are more stringent than leachate standards presently being proposed under Section 3001 of the Resource Conservation and Recovery Act (RCRA) by a factor of 10, underscoring the conservative approach taken by Westinghouse.

The only components of FBC leachate which consistently exceeded the drinking water standards were the following:

- Ca;
- SO_4;
- pH; and
- total dissolved solids.

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*Note, this set of experiments did not use the EP procedure (acetic acid) as described in the RCRA Guidelines published in the December 18, 1979 Federal Register. As mentioned earlier, tests done subsequent to these experiments using the acetic acid EPA procedure still showed no problems with FBC leachates when compared with the RCRA Guidelines.

TABLE 76. COMPARISON OF LEACHATE CHARACTERISTICS FROM THE FBC AND FGD RESIDUES[14]

Substance	Liquor (mg/l) FGD	Leachate (mg/l)		Drinking Water[*] Standards (mg/l)
		FBC	FGD	
Al	0 to 20	0 to >2	<1	
Ag	<0.05	<0.05	<0.05	0.05
As	<0.05	<0.05	0 to 0.1[‡]	0.05
B	>5	0 to >5	>1	
Ba	<1	<2[†‡]	<1	1.0
Be	<0.02	<0.02	<0.02	
Bi	<0.04	<0.04	<0.04	
Ca	>500[‡]	>500[‡]	>500[‡]	75
Cd	0 to 0.2[‡]	<0.01	<0.01	0.01
Co	<0.1	<0.1	<0.1	
Cr	<0.05	<1.0[*‡]	<0.05	0.05
Cu	<1	<0.1	<0.1	1.0
Fe	<0.3	<0.3	<0.3	0.3
Hg	<0.002	<0.002	<0.002	0.002
Mg	0 to >1,000[‡]	0 - 250[†‡]	0 to 500[‡]	50
Mn	0 to 20[‡]	<0.05	0 to 0.1[‡]	0.05
Mo	0.1 to 7.0	<5	<1	
Na	0 to >100	0 to >100	<10	
Ni	<1	<0.1	<0.1	2.0
Pb	<0.05	<0.05	<0.05	0.05
Sb	<0.2	<0.1	<0.1	
Se	0.001 to 0.5[‡]	<0.01	0 to 0.1[‡]	0.01
Si	0 to 30	0 to 30	0 to 5	
Sn	<1.0	<1.0	<1.0	1.0
Sr	0 to 40	0 to >10	0 to 5	
Ti	<2	<2	<2	
V	<2	<1	<1	
Zn	<2	<1	<1	5.0
Zr	<2	<1	<1	
SO_3	<10 to 40	<10	<10	
SO_4	1,000 to 7,000[‡]	1,000 - 2,000[‡]	1,000 - 2,000[‡]	250
Cl	300 to 6,000[‡]	0 to 350[†‡]	30 to 300[‡]	250
F	10 to 50[‡]	<2.4	1 to 10[‡]	2.4
NO_3 (as N)	0 to 100[‡]	<10	<10	10
TOC	<30	<30	<30	
pH	6 to 9	9 to 12[‡]	6 to 9	5 to 9
TDS	5,000 to 14,000[‡]	2,000 to 4,000[‡]	2,000 to 3,000[‡]	500
Specific Conductance millimhos/cm	5.0 to 17.0	0.5 to 10.0	2.0 to 3.0	

[*] National Interim Primary Drinking Water Regulations (NIPDWR) (1975) and U.S. Public Health Service (USPHS) (1962) Drinking Water Standards, and World Health Organization (WHO) Potable Water Standards.

[†] Concentrations higher than the Drinking Water Standards resulted from leachates of <2 Batches of carry-over fines among the >30 spent FBC materials investigated.

[‡] Exceed Drinking Water Standards.

In addition, the following species exceeded the drinking water standards in less than two of the more than 30 FBC samples tested:

- Ba;
- Cr;
- Mg; and,
- Cl.

Westinghouse[16] has concluded from their study that:

- No water pollution is expected from the leaching of those trace-metal ions for which drinking water standards exist, since the leachate itself meets drinking water standards.
- An insignificant amount of magnesium is leached out, even for dolomite sorbent.
- Sulfide may not be a problem for the once-through sorbent, since the sulfide concentration in the leachate is below detection limits.
- The total dissolved organics are below detection limits.
- Residual activity, reflected by heat release upon initial exposure to water, has been observed for once-through atmospheric pressure FBC systems, and is judged as an environmental concern for direct disposal. The heat release is attributed to the large amount of calcium oxide present in the spent sorbent.
- Potential problems with the leachates are the high concentrations of calcium (Ca), sulfate (SO_4), pH, and total dissolved solids (TDS), which are above drinking water standards.
- The addition of 20 wt percent ash to the spent sorbent improves leachate quality. Thus, codisposal of spent sorbent and ash can reduce the adverse environmental impact.
- The environmental impact is reduced by room-temperature processing.

According to Westinghouse, FBC residue will not be a hazardous pollutant; however, it is still a candidate for the RCRA special waste category by virtue of the volume of material which will be produced.

Further engineering and experimental studies are required in order to further define the environmental impact of the FBC residue in the actual disposal environment, and to systematically assess the design, performance and costs of alternative handling and disposal options. The following areas need investigation:

1. Define environmental impact of disposal.

   A more comprehensive view of the environmental impact of FBC residue can be approached through development of a methodology to project the environmental impact of commercial-scale disposal sites based upon laboratory data. Specifically, soil attenuation and deattenuation studies; field cell work (to assess the tendency of the material to set up, and the leaching properties that result); and further confirmation of the major environmental problems (pH, TDS, Ca, SO_4) on a wider variety of FBC residues should be pursued.

2. Assess handling options.

   Handling options for the solids prior to disposal must be identified and evaluated. Two options are hydration of the solid waste piles at the FBC plant site, or transporting the waste to a disposal site prior to hydration in covered trucks to avoid fugitive emissions during transport. These and other options and their environmental impacts must be assessed.

3. Assess disposal options.

   Disposal options must be identified and evaluated more fully. Options such as solid waste neutralization to control pH, clay-lined basins to prevent leaching to ground water, and pretreatment as a cement-like material at the site to prevent heat release and leaching, should be considered. Tests which are needed to evaluate these methods must also be identified. For example, liners must be tested to see whether the waste will react with the material or not, and what the consequences of any such reaction might be. The effectiveness of the liners must be assessed as well as any pretreatment options.

   Furthermore, it would be advantageous to the commercialization of FBC systems to follow the development of RCRA requirements as well as any other regulatory activities which may affect the disposal requirements for FBC, such as effluent guidelines or ground water regulations which may be developed in the future. The assessment of the cost of meeting these kinds of requirements is also essential.

It appears at this time that FBC solid waste disposal should not be an insurmountable problem. However, attention to suitable handling and disposal options should be given in the FBC plant design and cost studies.

6.2.2.3 Means of Reducing the Quantity of Solid Waste Generated--

Due to the large amount of solid waste generated, it is to the FBC developer and operator's best interest to reduce the quantity generated by whatever means are available. Methods of lowering the volume of material that are presently feasible are:

- using low sulfur coal;
- using a sorbent with high reactivity;
- increasing gas residence time; and,
- decreasing sorbent particle size.

Methods which are presently under investigation and development are:

- other methods of improving calcium utilization, such as injection of sodium chloride or calcium chloride;
- spent stone regeneration;
- alternate synthetic sorbents which require less volume and have better regeneration qualities; and,
- reactivation of spent stone by exposure to water.

6.2.2.4 Comparison of FBC Solid Waste with FGD Sludge--

The solid waste produced at an industrial-sized AFBC plant may range from 110 to 3,900 kg/hr (250 to 8,500 lb/hr). Babcock and Wilcox Company compared the solid waste from a fluidized-bed boiler with particulate control and from a flue gas desulfurization (FGD) system plus a precipitator on a conventional pulverized fuel boiler, using a 3 percent sulfur, 10 percent ash coal. Table 77 indicates the relative amount of material to be disposed of from the two systems.[17]

TABLE 77. BABCOCK AND WILCOX COMPANY'S COMPARISON OF SOLID WASTE MASS FROM FBC AND FGD[17]

	Quantity/ton of coal, kg (lbs)	
	FBC	FGD and precipitator
Ca/S Ratio required	4.0	1.1
$CaCO_3$ Supplied	349 (750)	94 (206)
Spent Sorbent	243 (536)	107 (235)
Limestone inert	18 (39)	5 (11)
Moisture in filter cake at 50 percent	-0-	112 (246)
Fly ash and carbon	105 (232)	105 (232)
Solid waste to haul away	366 (807)	329 (724)*
Form of Waste	Dry granular solid	Wet sludge

*Wet basis.

Based on the Babcock and Wilcox results, the mass of waste which must be hauled away to a landfill from an FBC boiler (dry basis) is only 11 percent more than the waste from a wet scrubber system (wet basis) when a separate (dry) particulate removal system is used with the scrubber. However, if the fly ash is also collected in the wet scrubber, then the amount of wet sludge will be greater than the amount of dry waste from the FBC boiler, due to the moisture content that would be associated with the fly ash in such a case.

The Tennessee Valley Authority has also compiled information on the relative mass of the two wastes produced. Table 78 indicates that although the actual amount of dry sorbent used is less for FGD than FBC, the solid waste mass is actually greater by as much as 40 percent due to the water content of the slurry.[18]

There are a few major differences between the waste from FBC and lime/limestone FGD. Listed below are the major environmental concerns associated with the waste from the two technologies.

FBC	FGD[19]
pH	pH
TDS	TDS
-	SO_3
SO_4	SO_4
Ca	Ca
-	Cl
dry granular solid	thixotropic sludge
heat release	-

TABLE 78. COMPARISON OF AFBC AND SCRUBBER SOLID WASTES FOR A 200 MW PLANT ESTIMATED BY TVA[18]

	AFBC	Conventional with scrubber
Coal		
Ash	10%	10%
Sulfur	3.5%	3.5%
Percent removal	85%	85%
Ca/S	2.5	1.5
Annual coal use	450,000 ton/yr	450,000 ton/yr
Spent Sorbent		
Dry	120,000 ton/yr	*
Wet	-	168,000 ton/yr
Spent Ash	45,000 ton/yr	45,000 ton/yr
Total waste	165,000 ton/yr	213,000 ton/yr

*84,000 ton/yr

The presence of sulfite ion ($SO_3^=$) in the scrubber sludge is the major chemical difference between the wastes produced by the two systems. From an environmental viewpoint this makes scrubber sludge a detriment, as this $SO_3^=$ will be a source of chemical oxygen demand, since it is readily oxidized to $SO_4^=$.

FGD sludge is a thixotropic, partially oxidized slurry. Since thixotropic slurry tends to liquefy easily, it is difficult to handle, and dewatering techniques such as centrifuges and vacuum filters do not reliably yield the 70 to 75 percent solids needed prior to landfilling.

FBC waste in contrast is a dry, almost fully oxidized solid, although in some cases it may be necessary to wet it down for handling purposes. It would not, however, contain as much water as scrubber sludge. The preliminary environmental concern with FBC waste is the leachate quality and heat release properties. Although the disposal of waste from the FBC system may be less of an environmental detriment than that from FGD, there is still a great volume of material which must be disposed of. Methods of lowering the volume of material that are presently feasible are using a low sulfur coal or a sorbent with high reactivity. Methods which are presently under investigation and development are:

- methods for improved calcium utilization;
- methods of regenerating spent stone; and,
- alternate synthetic sorbents which require less volume and better regeneration qualities.

Both residues may need some sort of treatment prior to disposal: FBC to control the heat release potential and FGD to dewater and oxidize the waste. It is difficult at this time to project exactly what degree of treatment will be necessary for either waste. According to the TVA and B&W studies, FBC waste has a slight disposal cost advantage over FGD sludge. Further study of this issue is warranted.

6.2.2.5 Byproduct Uses for Solid Waste--

The potential of this waste material as a byproduct should not be ignored. Because of the high amount of unused lime (CaO), uses as a cement supplement, agricultural additive, building material and road aggregate have all been explored and results are promising. As larger quantities of waste become available from the operation of a demonstration plant, a better assessment of the resource recovery possibilities can be made.

The Department of Energy (DOE) is funding a 5-year research program to identify and evaluate potential agricultural applications for FBC solid wastes.[20] The study is being performed simultaneously in several states, all located in the Eastern half of the United States. The program covers almost the entire crops grown in Eastern United States. It includes both short- and long-term laboratory and field based evaluations. The waste is used as a replacement for lime to neutralize soil, as a source for trace and certain nutrient elements, and as a source for sulfur. The study evaluates both the quality and quantity of crops produced from soil treated by waste material, as well as the crops' nutrient value as food for domestic animals.

A study to evaluate the physiological effects of food that was ultimately obtained from FBC waste-treated soils on people and animals has been proposed to DOE and EPA. The study will monitor mineral balance and amino acids in human tissues, primarily human hairs, which tend to accumulate toxic materials. Some small animals will be evaluated over several reproductive cycles to determine long term effects on offspring. People will be fed in two stages. The first test will start in October 1979 and the second is scheduled for 1980.

Several other studies have demonstrated that FBC solid residues, because of their unique chemical composition, possess cementitious characteristics which, if exploited, can turn the waste into a very durable concrete-like mass. One such study investigated the potential for using FBC solid waste for road constructions.[21] The result indicated that comprehensive strength of cemented waste exceeded the value recommended for heavy traffic highway construction over a wide range of compositions. Further, this compressive strength, which is indicative of the material durability and resistance to erosion, improved with time even after the cemented samples were subjected to the effect of freeze/thaw cycles. The study concluded that the exceptionally high strength of cemented FBC residue makes it suitable for applications requiring materials with low water permeability, such as in embankments, structural fills, and liners to control leaching from waste disposal dumps and lagoons. The latter application is particularly important, since some clay-type liners which are being used in sanitary landfill have developed cracks after several years of use, which allow leachates to further percolate into ground water aquifers.

6.2.3 Water Pollution

Most aqueous emissions from AFBC such as boiler feed water treatment effluents, thermal discharge, and runoff from coal and limestone piles will be similar to conventional boilers' effluents.

Water pollution from solid waste disposal is discussed in Section 6.2.2. The preliminary water impact concerns are the pH, TDS, Ca, and SO_4 contents of the leachate.

New FBC sites will be required to obtain a National Pollution Discharge Elimination System (NPDES) permit under the Water Pollution Control Act requiring zero discharge (not increasing the pollution level of waters) of certain pollutants such as TDS, pH, BOD (biochemical oxygen demand) and COD (chemical oxygen demand), and certain other pollutants which are characteristic of the particular process (i.e., possibly SO_4 and Ca for FBC).

6.3 OIL-FIRED AFBC

The air, water, and solid waste pollution from oil-fired AFBC units is expected to be similar to coal-fired pollutants. It is expected that air emissions will be lower due to lower fuel sulfur content, lower nitrogen content, and lower fuel ash. The solid waste impact, therefore, will also be lower because less sorbent feed would be required to remove the SO_2.

6.4 SUMMARY

6.4.1 Impact of Emission Control Technique

In terms of implementing the best candidates for emission control in fluidized-bed combustion, the major environmental concern is the impact of increased Ca/S mole ratios for SO_2 control on the amount of solid waste generated. Enhanced SO_2 control using high Ca/S ratios along with very small limestone particle sizes could also increase particulate emissions, but it is doubtful that this increase would be to such a degree that available particle control systems could not handle it.

Implementing the three levels of NO_x control requires little change in operating variables and little if any environmental impact is foreseen.

The only major environmental impact foreseen in implementing moderate, intermediate or stringent particulate control is the concomitant 5 to 10 percent increase in solid waste disposal associated with the increased control. Characterization of the nature of these collected fines is an area where further research is needed.

6.4.2 Solid Waste Disposal

FBC residue does not currently appear to be "hazardous" under RCRA Section 3001 (i.e., it is not considered toxic, reactive, corrosive, or ignitable). However, future RCRA developments do need to be followed.

Potential problems associated with the residue which have been identified are: the pH, TDS, Ca and SO_4 in the leachate and initial heat release upon contact with water and total solid mass, and handling problems.

Thus, the residue will require some care in handling and disposal such as pretreatment with water, neutralization, clay-lined basins for disposal, or a combination of these options. Generally, the disposal of AFBC residue does not pose any insurmountable problems.

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7.0 EMISSION SOURCE TEST DATA

7.1 INTRODUCTION

Most of the emission test data from coal-fired atmospheric FBC boilers has been obtained using sampling and analytical techniques other than EPA reference methods. A variety of circumstances have contributed to this fact. Primarily, emission data have been collected on experimental units, mostly to characterize the FBC process and to investigate emission variability as a function of boiler operating conditions. Because of the experimental nature of many of the early units, and because the emissions from such small units may not be completely characteristic of emissions from full-scale commercial units, rigorous testing to determine compliance with specific emission standards had not been an issue. In addition, due to the fact that previously available FBC units were generally not amenable to continuous (24 hr/day) long-term operation for extended periods, no long-term averaging (e.g., 30 day periods) data have been generated. Most test periods have been short, some only hours, others a few days. Although various investigators have expressed emission results in terms of emission standards, EPA reference methods were not always rigorously followed; in fact, a significant portion of the available emissions data from atmospheric FBC units was obtained before the EPA reference methods were officially accepted.

Sampling and analysis techniques have varied widely depending on the needs and equipment limitations of individual experimental programs. The nature of some of the FBC pilot facilities made certain aspects of compliance testing impractical; i.e., traversing across very small ducts and locating sampling ports at stipulated distances from upstream and downstream disturbances. In addition, there has been no impetus for monitoring according to EPA reference methods in FBC testing performed in foreign countries, such as England, where a significant portion of the early work was done.

A high priority in current testing plans is to monitor large scale FBC boiler facilities for 30 day periods using EPA reference methods. Planning is underway now to conduct such testing as soon as appropriate large AFBC units begin operating for extended periods.

This section emphasizes the results of test programs conducted at the largest atmospheric FBC units. Much of these data were obtained several years ago, and in many cases, important design conditions, such as gas phase residence time, freeboard height, and limestone particle size were not necessarily optimal.* The larger FBC units discussed here include those operated by Babcock and Wilcox (B&W); the National Coal Board (NCB); Pope, Evans and Robbins (PER); and Babcock and Wilcox, Ltd. (Great Britain).

This section also reports raw test data which was referred to or summarized in Section 2.0 of this report. A large portion of the discussion and results presented in Section 2.0 is based on testing results reported by NCB, PER, and B&W.

A general description of the FBC test facilities noted in this section is presented in Table 79. A more detailed description of each test facility

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*As discussed elsewhere in this report, recent theoretical and bench scale experimental work indicate substantial increases in SO_2 removal efficiency can result using longer residence times and smaller limestone particles.

TABLE 79. GENERAL DESCRIPTION OF ATMOSPHERIC FBC TEST FACILITIES

Investigator	FBC unit designation	Boiler classification and capacity (as tested)	Bed depth-meters (ft)	Fluidizing velocity m/sec (ft/sec)	Other design features	Emission data reported*	Remarks	Reference number
Babcock and Wilcox (B&W)	6 ft × 6 ft	Pilot scale 7 MW_t (25 × 10^6 Btu/hr)	0.80 - 1.44 (2.83 - 4.73)	3 (8)	System includes primary cyclone	SO_2, NO_x particulate	Demonstrates greater than 90 percent SO_2 control. Limited recycle possible (only 25 percent of carryover)	1 and 2, 3
Babcock and Wilcox (B&W)	3 ft × 3 ft	Pilot scale 1.9 MW_t (6.5 × 10^6 Btu/hr)	0.3 - 0.6 (1.0 - 2.1)	1.2 - 3.7 (4 - 12)	Integral water jacketed fly ash removal device	SO_2, NO_x particulate	Shallow bed design not optimal for SO_2 reduction. Low freeboard, no recycle	4
National Coal Board (NCB)	3 ft × 1.5 ft	Pilot scale 0.3 - 1.3 MW_t (1 - 4.5 × 10^6 Btu/hr)	0.6 - 2.1 (2.0 - 7.0)	0.6 - 2.4 (2.0 - 8.0)	System includes primary and secondary cyclones	SO_2, NO_x	Underbed bed feed with recycle. High freeboard.	5, 6
Pope, Evans, and Robbins	1.5 × 6 ft	Full scale boiler module 3.2 MW_t (11 × 10^6 Btu/hr)	0.3 - 0.6 (1.0 - 2.0)†	3.0 - 4.6 (10 - 15)	Integral multiclone collector for primary fly ash removal	SO_2, NO_x particulate	Short gas phase residence time and shallow bed depth not optimal for SO_2 removal†	7, 8
Babcock and Wilcox, Ltd.	10 ft × 10 ft	Industrial scale 12 MW_t (40 × 10^6 Btu/hr)	-	-	FBC retrofit to stoker-fired boiler	SO_2, NO_x	Details of boiler design and test procedures are not currently available	9
FluiDyne	3 ft × 5.3 ft	Pilot scale 68 - 227 kg/hr coal	1.1 - 1.2 (3.5 - 4.0)	0.6 - 1.3 (2.0 - 4.2)	Underbed or overbed feed with recycle	SO_2	Effective SO_2 control due to long gas phase residence time	10, 11
FluiDyne	1.5 ft × 1.5 ft	Pilot scale	-	-	Underbed or overbed feed with recycle	SO_2	Demonstrated equivalent desulfurization with inbed or overbed feed and primary recycle	10
National Coal Board (NCB)	6 in. diameter	Bench scale	0.6 - 0.9 (2.0 - 3.0)	0.6 - 0.9 (2.0 - 3.0)	Underbed feed with recycle	SO_2	Effective SO_2 control consistent with operating conditions close to those recommended for "best system" design	12
Argonne National Laboratories (ANL)	6-in. diameter	Bench scale	0.38 - 0.61 (1.25 - 2.0)	0.73-2.36 (2.40-7.73)	Underbed feed with recycle	SO_2, NO_x	Effective SO_2 control consistent with operating conditions close to those recommended for "best system" design. NO_x higher than expected due to small unit size.	13, 14, 15 and 16

*In some cases, more data was originally reported, but only emissions pertinent to this investigation are tabulated in this section.

†Static bed depth.

is presented in Subsection 7.4. Emissions measured that are of concern in this effort are also presented in Table 79.

7.2 EMISSION SOURCE TEST DATA FOR COAL-FIRED ATMOSPHERIC FBC BOILERS

This subsection presents detailed raw test data for experimental AFBC test units. Table 80 is an index of the investigators and test units for which data is reported.

TABLE 80. INDEX OF AFBC EMISSION TEST DATA

Table No.	Investigator	FBC unit designation	Year of testing
81	B&W	6 ft × 6 ft	1978-1979
82	B&W	3 ft × 3 ft	1976
83	NCB	CRE 3 ft × 1.5 ft	1970-1971
84	PER	FBM 1.5 ft × 6 ft	Late 1967 through 1969
85	PER	FBM 1.5 ft × 6 ft	Through 1975
86	FluiDyne	Vertical slice FBC (3.3 ft × 5.3 ft)	1977
87	FluiDyne	Vertical slice FBC (3.3 ft × 5.3 ft)	1977
88	NCB	6 in. diameter	1970-1971
89	NCB	6 in. diameter	1970-1971
90	NCB	6 in. diameter	1970-1971
91	Argonne	6 in. diameter	1968-1971

In addition, graphical emissions data reported by B&W, Ltd. at the Renfrew, Scotland boiler are included in Figures 57 and 58. Figure 59 is a graph of data recorded by FluiDyne during operation of their 1.5 ft × 1.5 ft unit.

TABLE 81. EMISSION TEST DATA MEASURED FROM B&W 6 FT × 6 FT AFBC UNIT FIRING OHIO NO. 6 COAL WITH LOWELLVILLE LIMESTONE, SIZED ≤9510 μm (3/8 in. × 0)[1,2,3]

Test No.	Date	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas Residence time sec	Fuel characteristics				Sorbent characteristics		Emissions characteristics						
												SO_2 furnace exit[*]			Dust loading (cyclone inlet)[†]		Dust loading (cyclone outlet)[‡]	
						Heating value kJ/kg (Btu/lb)	% S	% Ash	Feed rate g/s (lb/hr)	Feed rate g/s (lb/hr)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	% SO_2 retention	g/s (lb/hr)	ng/J (lb/10^6 Btu)	g/s (lb/hr)	ng/J (lb/10^6 Btu)
1-1	4-30-78	876 (1,609)	2.42 (7.95)	1.13 (46.6)	0.47	28,907 (12,436)	3.46	7.29	270.8 (2,149)	136.1 (1,080)	4.22	167	133.3 (0.31)	94.6	70.8 (562)	9,028 (21.0)	NR (NR)	NR (NR)
1-1	5-1-78	878 (1,612)	2.50 (8.21)	1.21 (47.6)	0.68	28,907 (12,436)	3.46	7.29	270.8 (2,149)	136.1 (1,080)	4.22	174	133.3 (0.31)	94.3	70.8 (562)	9,028 (21.0)	NR (NR)	NR (NR)
1-2	5-2-78	865 (1,588)	2.33 (7.66)	1.42 (55.9)	0.61	31,217 (13,430)	3.27	6.83	259.4 (2,059)	137.5 (1,091)	4.80	90	64.5 (0.15)	96.8	83.3 (661)	11,092 (25.8)	NR (NR)	NR (NR)
1-2	5-2-78	869 (1,597)	2.55 (8.36)	1.42 (55.88)	0.56	31,217 (13,430)	3.27	6.83	259.4 (2,059)	137.5 (1,091)	4.80	90	68.8 (0.16)	96.8	83.3 (661)	10,275 (23.9)	NR (NR)	NR (NR)
1-2	5-2-78	871 (1,600)	2.38 (7.30)	1.44 (56.76)	0.61	31,242 (13,440)	3.28	6.84	270.1 (2,144)	137.5 (1,091)	4.51	131	94.6 (0.22)	95.5	83.3 (661)	9,888 (23.0)	NR (NR)	NR (NR)
1-2	5-2-78	874 (1,605)	2.50 (8.19)	1.42 (56.07)	0.57	31,242 (13,440)	3.28	6.84	270.1 (2,144)	137.5 (1,091)	4.51	126	94.6 (0.22)	95.7	83.3 (661)	9,888 (23.0)	NR (NR)	NR (NR)
1-3	5-4-78	872 (1,601)	2.56 (8.39)	1.24 (48.96)	0.48	28,970 (12,464)	3.20	8.18	258.6 (2,052)	125.5 (996)	4.59	138	111.8 (0.26)	95.1	49.0 (389)	6,635 (15.2)	NR (NR)	NR (NR)
1-3	5-4-78	867 (1,592)	2.53 (8.29)	1.23 (48.26)	0.49	30,464 (13,106)	3.20	8.18	258.6 (2,052)	125.5 (996)	4.06	135	107.5 (0.25)	95.2	49.0 (389)	6.635 (15.2)	NR (NR)	NR (NR)
1-3	5-4-78	867 (1,592)	2.48 (8.16)	1.27 (49.86)	0,51	30,464 (13,106)	3.47	8.82	258.6 (2,052)	125.5 (996)	4.59	135	133.3 (0.31)	95.2	49.0 (389)	5,976 (13.9)	NR (NR)	NR (NR)
1-4	5-6-78	849 (1,559)	2.06 (6.75)	0.85 (33.33)	0.41	31,589 (13,590)	3.39	5.93	193.5 (1,536)	102.1 (810)	4.50	156	116.1 (0.27)	94.2	49.2 (390)	8,040 (18.7)	NR (NR)	NR (NR)
1-4	5-6-78	852 (1,565)	2.08 (6.87)	0.85 (33.33)	0.41	31,589 (13,590)	3.39	5.93	193.5 (1,536)	102.1 (810)	4.50	163	120.4 (0.28)	94.0	49.2 (390)	8,040 (18.7)	NR (NR)	NR (NR)
1-4	5-6-78	866.3 (1,591)	2.13 (6.98)	0.80 (31.67)	0.38	31,589 (13,590)	3.29	6.83	192.8 (1,530)	96.4 (765)	4.46	140	107.5 (0.25)	94.6	49.2 (390)	8,169 (19.0)	NR (NR)	NR (NR)
1-4	5-6-78	855.6 (1,572)	2.06 (6.74)	0.82 (32.43)	0.40	31,589 (13,590)	3.28	6.83	192.8 (1,530)	96.4 (765)	4.46	130	98.9 (0.23)	94.9	49.2 (390)	8,169 (19.0)	NR (NR)	NR (NR)
1-5	5-7-78	872 (1,601)	2.49 (8.15)	1.19 (47.03)	0.48	31,426 (13,520)	3.14	6.28	246.7 (1,958)	117.8 (935)	4.20	183	133.3 (0.31)	93.3	54.8 (435)	7,057 (16.4)	NR (NR)	NR (NR)
2-1	6-8-78	869 (1,596)	2.56 (8.39)	1.19 (47.0)	0.46	29,506 (12,694)	3.48	6.68	263.3 (2,090)	81.9 (650)	2.69	700	541.8 (1.26)	78.8	69.9 (555)	8,985 (20.9)	25.7 (170)	2,750 (6.4)
2-2	6-10-78	839 (1,542)	2.45 (8.02)	1.26 (49.53)	0.51	29,209 (12,566)	4.05	8.10	254.5 (2,020)	84.4 (670)	2.44	891	688 (1.60)	77.3	64.4 (511)	8,641 (20.1)	24.0 (159)	2,710 (6.3)
2-2	6-10-78	845 (1,553)	2.32 (7.61)	1.25 (49.07)	0.56	29,426 (12,660)	3.96	7.25	245.7 (1,950)	89.5 (710)	2.78	588	464 (1.08)	85.4	68.6 (543)	9,458 (22.0)	24.6 (163)	2,840 (6.6)
2-2	6-10-78	844 (1,551)	2.32 (7.61)	1.26 (49.5)	0.54	29,426 (12,660)	3.96	7.25	245.7 (1,950)	89.5 (710)	2.78	584	443 (1.03)	86.0	68.6 (543)	9,458 (22.0)	24.6 (163)	2,840 (6.6)
2-2	6-10-78	845 (1,553)	2.40 (7.88)	1.26 (49.6)	0.53	29,426 (12,660)	3.96	7.25	245.7 (1,950)	89.5 (710)	2.78	604	473 (1.10)	84.4	68.4 (543)	9,458 (22.0)	24.6 (163)	2,840 (6.6)
2-2	6-10-78	842 (1,548)	2.46 (8.06)	1.21 (47.75)	0.49	29,492 (12,690)	3.88	7.01	253.3 (2,010)	103.3 (820)	3.20	774	602 (1.40)	79.8	77.1 (612)	10,318 (24.0)	38.1 (252)	4,260 (9.9)
2-2	6-10-78	846 (1,556)	2.40 (7.86)	1.20 (47.19)	0.50	29,188 (12,556)	3.75	6.84	252.0 (2,000)	93.24 (740)	2.95	676	529 (1.23)	81.9	67.7 (537)	9,200 (21.4)	24.4 (162)	2,790 (6.5)
2-2	6-10-78	845 (1,554)	2.32 (7.60)	1.23 (48.57)	0.53	29,188 (12,556)	3.75	6.84	252.0 (2,000)	93.24 (740)	2.95	681	533.3 (1.24)	82.3	67.7 (537)	9,200 (21.4)	24.4 (162)	2,790 (6.5)
2-2	6-10-78	845 (1,553)	2.43 (7.98)	1.23 (48.25)	0.51	29,188 (12,556)	3.75	6.84	252.0 (2,000)	93.24 (740)	2.95	627	494.5 (1.15)	82.8	67.7 (537)	92200 (21.4)	24.4 (162)	2,790 (6.5)
2-3	6-15-78	841 (1,546)	2.17 (7.11)	1.22 (48.22)	0.56	29,784 (12,838)	3.21	6.32	244.4 (1,940)	0 (0)	0	1,495	1,152.4 (2.68)	55.0[‡]	36.1 (271)	4,686 (10.9)	16.3 (108)	1,850 (4.3)
2-3	6-15-78	841 (1,546)	2.17 (7.11)	1.22 (48.22)	0.56	29,784 (12,838)	3.21	6.32	244.6 (1,940)	0 (0)	0	1,495	1,152.4 (2.68)	55.0[‡]	34.1 (271)	4,686 (10.9)	16.3 (108)	1,850 (4.3)

TABLE 81 (continued)

[illegible]	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas residence time sec	Fuel characteristics: Heating value kJ/kg (Btu/lb)	% S	% N	% Ash	Feed rate g/s (lb/hr)	Sorbent characteristics: Feed rate g/s (lb/hr)	Ca/S ratio	Emission characteristics: SO_2 ppm	SO_2 ng/J (lb/10^6 Btu)	SO_2 retention	NO_x ppm	NO_x ng/J (lb/10^6 Btu)	Particulate: Cyclone inlet ng/J (lb/10^6 Btu)	Particulate: Cyclone outlet ng/J (lb/10^6 Btu)
5-2A	838 (1,541)	2.69 (8.8)	1.21 (47.7)	0.45	29,508 (12,686)	4.34	1.23	6.62	248 (1,965)	72 (575)	1.87	962	770 (1.79)	76.30	-	-	18,904 (43.97)	3,224 (7.50)
5-2B	841 (1,545)	2.63 (8.62)	1.22 (48.2)	0.46	29,508 (12,686)	4.34	1.23	6.62	248 (1,965)	77 (610)	1.98	1,041	838 (1.95)	75.03	-	-	18,904 (43.97)	3,224 (7.50)
6-1A	847 (1,556)	2.5 (8.3)	1.2 (47.7)	0.48	29,894 (12,852)	3.69	1.13	6.05	243 (1,928)	81 (640)	2.64	785	580 (1.35)	76.40	-	-	12,451 (28.86)	6,214 (14.57)
6-1B	847 (1,556)	2.5 (8.2)	1.2 (47.6)	0.48	29,894 (12,852)	3.69	1.13	6.05	242 (1,921)	81 (640)	2.65	785	572 (1.33)	76.79	-	-	12,451 (28.96)	6,285 (14.62)
6-1C	849 (1,560)	2.7 (8.8)	1.1 (45.2)	0.41	29,894 (12,852)	3.69	1.13	6.05	243 (1,932)	81 (640)	2.63	751	589 (1.37)	76.17	-	-	12,382 (28.80)	6,243 (14.54)
6-1D	847 (1,557)	2.5 (8.3)	1.2 (47.6)	0.48	29,722 (12,778)	3.77	1.13	6.13	242 (1,918)	81 (640)	2.63	771	576 (1.34)	77.35	-	-	12,541 (29.17)	6,333 (14.73)
6-1E	847 (1,556)	2.6 (8.6)	1.2 (47.9)	0.46	29,722 (12,778)	3.77	1.13	6.13	246 (1,956)	81 (640)	2.58	749	592 (1.33)	77.38	-	-	12,300 (28.61)	6,208 (14.44)
6-1F	848 (1,558)	2.6 (8.5)	1.2 (47.4)	0.46	29,740 (12,786)	3.69	1.13	6.24	243 (1,930)	81 (640)	2.65	806	615 (1.43)	75.17	-	-	12,455 (28.97)	6,290 (14.63)
6-1G	846 (1,555)	2.6 (8.5)	1.2 (46.8)	0.46	29,740 (12,786)	3.69	1.13	6.24	245 (1,947)	81 (640)	2.63	826	623 (1.45)	74.83	-	-	12,347 (28.72)	6,234 (14.50)
6-1H	846 (1,554)	2.6 (8.5)	1.2 (48.0)	0.46	29,740 (12,786)	3.69	1.13	6.24	243 (1,926)	81 (640)	2.66	713	542 (1.26)	78.07	-	-	12,481 (29.03)	6,303 (14.66)
6-1I	851 (1,563)	2.6 (8.6)	1.2 (46.2)	0.46	29,663 (12,753)	3.87	1.23	6.32	238 (1,886)	81 (640)	2.61	706	559 (1.30)	78.54	-	-	12,782 (29.73)	6,453 (15.01)
6-1J	850 (1,563)	2.6 (8.6)	1.2 (46.6)	0.46	29,663 (12,753)	3.87	1.23	6.32	240 (1,905)	81 (640)	2.38	748	585 (1.36)	77.60	-	-	12,653 (29.43)	6,389 (14.86)
6-1K	848 (1,559)	2.5 (8.3)	1.2 (47.1)	0.48	29,663 (12,753)	3.87	1.23	6.32	247 (1,961)	81 (640)	2.51	767	563 (1.31)	78.37	-	-	12,291 (28.59)	6,210 (14.44)
6-1L	837 (1,539)	2.6 (8.5)	1.2 (49.0)	0.46	29,117 (12,518)	3.65	1.22	7.50	227 (1,801)	68 (540)	2.47	917	770 (1.79)	69.36	-	-	9,709 (22.58)	3,341 (7.98)
6-1M	834 (1,534)	2.5 (8.3)	1.1 (44.5)	0.52	29,117 (12,518)	3.65	1.22	7.50	229 (1,818)	68 (540)	2.45	982	791 (1.84)	68.36	-	-	9,617 (22.37)	3,401 (7.91)
6-1N	835 (1,535)	2.6 (8.5)	1.2 (48.0)	0.46	29,117 (12,518)	3.65	1.22	7.50	233 (1,851)	68 (540)	2.40	1,070	877 (2.04)	65.06	-	-	9,445 (21.97)	3,340 (7.77)
6-1O	836 (1,537)	2.7 (8.9)	1.3 (49.9)	0.48	29,117 (12,518)	3.65	1.22	7.50	233 (1,850)	68 (540)	2.40	1,115	950 (2.21)	62.09	-	-	9,450 (21.98)	3,340 (7.77)
6-1P	834 (1,533)	2.6 (8.5)	1.2 (48.8)	0.46	29,117 (12,518)	3.65	1.22	7.50	234 (1,860)	68 (540)	2.39	1,067	860 (2.00)	65.68	-	-	9,398 (21.86)	3,323 (7.73)
6-2A	894 (1,641)	2.7 (8.8)	0.8 (32.1)	0.30	29,194 (12,551)	4.26	1.13	7.64	223 (1,773)	78 (620)	2.43	1,454	1,178 (2.74)	59.47	202	116 (0.27)	10,374 (24.13)	3,147 (7.32)
6-2B	897 (1,647)	2.6 (8.6)	0.8 (32.0)	0.30	29,194 (12,551	4.24	1.13	7.64	225 (1,783)	78 (620)	2.41	1,457	1,152 (2.68)	60.30	214	120 (0.28)	10,318 (24.00)	3,130 (7.28)
6-2C	896 (1,644)	2.6 (8.5)	0.8 (32.8)	0.30	29,194 (12,551)	4.24	1.13	7.64	225 (1,782)	78 (620)	2.42	1,502	1,169 (2.72)	59.65	198	112 (0.26)	10,322 (24.01)	3,134 (7.29)
6-3A	847 (1,557)	2.7 (8.8)	1.2 (47.5)	0.44	29,324 (12,607)	4.22	1.22	6.57	251 (1,990)	83 (660)	2.31	938	772 (1.68)	74.90	207	112 (0.26)	9,136 (21.25)	3,203 (7.45)
6-3B	847 (1,556)	[illegible]	1.2 (47.8)	0.50	29,324 (12,607)	4.22	1.22	6.57	238 (1,891)	83 (660)	2.43	836	598 (1.39)	79.20	204	107 (0.25)	9,613 (22.36)	3,371 (7.84)
6-3C	847 (1,556)	[illegible]	1.2 (47.0)	0.48	29,324 (12,607)	4.22	1.22	6.57	235 (1,866)	83 (660)	2.46	857	636 (1.48)	77.84	208	112 (0.26)	9,742 (22.66)	3,418 (7.95)

TABLE 81 (continued)

Test No.	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas residence time sec	Fuel characteristics					Sorbent characteristics		Emission characteristics						
												SO₂			NOₓ		Particulate	
																	Cyclone inlet	Cyclone outlet
					Heating value kJ/kg (Btu/lb)	% S	% N	% Ash	Feed rate g/s (lb/hr)	Feed rate g/s (lb/hr)	Ca/S ratio	ppm	ng/J (lb/10⁶ Btu)	SO₂ retention	ppm	ng/J (lb/10⁶ Btu)	ng/J (lb/10⁶ Btu)	ng/J (lb/10⁶ Btu)
4-3D	847 (1,558)	2.5 (8.3)	1.2 (48.3)	0.48	29,168 (12,540)	4.14	1.22	7.14	235 (1,868)	83 (660)	2.57	855	662 (1.54)	76.67	200	112 (0.26)	9,781 (22.75)	3,431 (7.98)
4-3E	849 (1,560)	2.5 (8.2)	1.2 (46.5)	0.48	29,168 (12,540)	4.14	1.22	7.14	236 (1,876)	83 (660)	2.56	818	615 (1.43)	78.32	218	77 (0.18)	9,742 (22.66)	3,418 (7.95)
5-1A	847 (1,556)	2.5 (8.2)	1.2 (47.7)	0.48	29,368 (12,626)	3.89	1.14	7.51	225 (1,784)	94 (750)	3.21	423	357 (0.83)	86.56	-	-	10,155 (23.62)	2,042 (4.75)
5-1B	851 (1,564)	2.6 (8.4)	1.2 (48.9	0.46	29,368 (12,626)	3.89	1.14	7.51	223 (1,768)	94 (750)	3.24	457	396 (0.92)	85.10	-	-	10,245 (23.83)	2,059 (4.79)
5-1C	852 (1,565)	2.6 (8.5)	1.2 (46.5)	0.46	29,368 (12,626)	3.89	1.14	7.51	225 (1,785)	94 (750)	3.21	557	477 (1.11)	81.93	241	150 (0.35)	10,150 (23.61)	2,042 (4.75)
5-1D	848 (1,558)	2.6 (8.5)	1.2 (47.2)	0.46	29,368 (12,626)	3.89	1.14	7.51	222 (1,762)	94 (750)	3.25	578	503 (1.17	81.00	254	159 (0.37)	10,275 (23.91)	2,068 (4.81)
5-1E	852 (1,565)	2.5 (8.3)	1.2 (47.6)	0.48	29,368 (12,626)	3.89	1.14	7.51	222 (1,765)	94 (750)	3.24	574	486 (1.13)	81.57	266	163 (0.38)	10,262 (23.87)	2,064 (4.80)
5-2A	843 (1,550)	2.6 (8.6)	1.2 (47.7)	0.46	28,987 (12,462)	3.94	1.22	7.31	229 (1,814)	101 (800)	3.35	574	477 (1.11)	82.36	306	185 (0.43)	10,060 (23.40)	1,367 (3.18)
5-2B	845 (1,554)	2.6 (8.6)	1.2 (47.1)	0.46	28,987 (12,462)	3.94	1.22	7.31	232 (1,842)	101 (800)	3.30	417	344 (0.80)	87.34	-	-	9,910 (23.05)	1,350 (3.14)
5-2C	861 (1,583)	2.6 (8.6)	1.2 (46.5)	0.46	28,987 (12,462)	3.94	1.22	7.31	234 (1,860)	101 (800)	3.27	577	456 (1.06)	83.23	-	-	9,811 (22.82)	1,337 (3.10)
5-2D	847 (1,557)	2.6 (8.6)	1.3 (49.8)	0.50	29,015 (12,474)	3.85	1.03	7.24	248 (1,970)	82 (650)	2.47	729	555 (1.29)	79.01	-	-	8,134 (18.92)	770 (1.79)
5-2E	836 (1,537)	2.6 (8.4)	1.2 (49.1)	0.46	29,015 (12,474)	3.85	1.03	7.24	233 (1,848)	113 (900)	3.64	766	610 (1.42)	77.01	-	-	8,672 (20.17)	821 (1.91)
5-2F	835 (1,534)	2.6 (8.4)	1.2 (47.2)	0.46	29,015 (12,474)	3.85	1.03	7.24	228 (1,813)	96 (765)	3.16	876	722 (1.68)	72.81	258	150 (0.35)	8,839 (20.56)	838 (1.95)
5-3A	898 (1,648)	2.9 (9.6)	1.2 (47.2)	0.41	28,791 (12,378)	4.12	1.12	7.68	256 (2,033)	84 (665)	2.38	979	808 (1.88)	71.78	-	-	9,329 (21.70)	2,205 (5.13)
5-3B	(1,646)	3.1 (10.0)	1.2 (46.8)	0.40	28,791 (12,378)	4.12	1.12	7.68	259 (2,058)	84 (665)	2.35	1,224	1,040 (2.42)	63.61	-	-	9,213 (21.43)	2,175 (5.06)
5-3C	899 (1,650)	3.0 (9.9)	1.2 (46.1)	0.40	28,791 (12,378)	4.12	1.12	7.68	265 (2,105)	78 (620)	2.14	1,073	881 (2.05)	69.25	-	-	9,011 (20.96)	2,128 (4.95)
6-1A	845 (1,552)	2.5 (8.1)	1.2 (48.8)	0.48	28,768 (12,368)	4.22	1.22	8.15	244 (1,940)	117 (926)	3.38	515	383 (0.89)	86.87	275	95 (0.22)	8,422 (19.59)	2,184 (5.08)
6-1B	850 (1,562)	2.5 (8.2)	1.2 (47.8)	0.48	28,768 (12,368)	4.22	1.22	8.15	246 (1,954)	108 (855)	3.10	479	361 (0.84)	87.72	275	99 (0.23)	8,362 (19.45)	2,171 (5.05)
6-1C	849 (1,560)	2.5 (8.1)	1.3 (50.1)	0.52	28,768 (12,368)	4.22	1.22	8.15	248 (1,971)	116 (918)	3.30	485	353 (0.82)	87.94	290	99 (0.23)	8,829 (19.28)	2,150 (5.00)
6-1D	849 (1,560)	2.5 (8.1)	1.2 (48.3)	0.48	28,768 (12,368)	4.22	1.22	8.15	245 (1,944)	68 (538)	1.97	482	361 (0.84)	87.72	290	103 (0.24)	8,405 (19.55)	2,180 (5.07)
6-1E	853 (1,567)	2.6 (8.6)	1.2 (47.3)	0.46	29,112 (12,516)	4.02	1.31	6.82	237 (1,879)	102 (813)	3.26	482	404 (0.94)	85.42	290	112 (0.26)	7,807 (18.16)	1,664 (3.87)
6-1F	853 (1,567)	2.6 (8.4)	1.2 (47.7	0.46	29,112 (12,516)	4.02	1.31	6 92	240 (1,906)	101 (804)	3.18	452	361 (0.84)	86.90	290	107 (0.25)	7,696 (17.90)	1,638 (3.81)
6-1G	851 (1,563)	2.6 (8.4)	1.2 (47.9)	0.46	29,112 (12,516)	4.02	1.31	6.82	238 (1,889)	100 (795)	3.17	520	426 (0.99)	84.36	300	116 (0.27)	8,057 (18.74)	1,948 (4.53)

TABLE 81 (continued)

Test No.	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas residence time sec	Fuel characteristics					Sorbent characteristics		Emission Characteristics						
												SO_2			NO_x		Particulate	
																	Cyclone inlet	Cyclone outlet
					Heating value kJ/kg (Btu/lb)	% S	% N	% Ash	Feed rate g/s (lb/hr)	Feed rate g/s (lb/hr)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention	ppm	ng/J (lb/10^6 Btu)	ng/J (lb/10^6 Btu)	ng/J (lb/10^6 Btu)
6-1H	849 (1,561)	2.6 (8.5)	1.2 (47.9)	0.46	29,112 (12,516)	4.02	1.31	6.82	238 (1,886)	101 (801)	3.20	514	426 (0.99)	84.59	300	116 (0.27)	8,070 (18.77)	1,948 (4.53)
6-1I	844 (1,551)	2.4 (8.0)	1.2 (48.7)	0.46	29,324 (12,607)	4.22	1.22	7.32	239 (1,896)	117 (931)	3.37	689	537 (1.25)	81.24	270	99 (0.23)	7,906 (18.39)	1,943 (4.52)
6-1J	845 (1,552)	2.5 (8.0)	1.2 (48.3)	0.48	29,324 (12,607)	4.22	1.22	7.32	240 (1,908)	117 (926)	3.33	649	507 (1.18)	82.35	270	99 (0.23)	7,846 (18.25)	1,930 (4.49)
6-1K	846 (1,554)	2.5 (8.1)	1.2 (48.6)	0.48	29,324 (12,607)	4.22	1.22	7.32	242 (1,929)	115 (915)	3.25	749	580 (1.35)	79.77	270	99 (0.23)	7,825 (18.20)	1,926 (4.48)
6-1L	841 (1,547)	2.3 (7.5)	1.2 (47.6)	0.52	29,810 (12,816)	3.25	1.24	8.02	224 (1,774)	64 (508)	2.68	608	469 (1.09)	78.57	285	103 (0.24)	6,561 (15.26)	1,702 (3.96)
6-1M	842 (1,548)	2.3 (7.4)	1.2 (48.9)	0.52	29,810 (12,816)	3.25	1.24	8.02	222 (1,759)	61 (486)	2.58	552	421 (0.98)	80.59	285	103 (0.24)	6,616 (15.39)	1,715 (3.99)
6-1N	842 (1,548)	2.3 (7.6)	1.2 (47.8)	0.52	29,810 (12,816)	3.25	1.24	8.02	224 (1,774)	61 (483)	2.55	574	447 (1.04)	79.56	285	103 (0.24)	6,561 (15.26)	1,702 (3.96)
6-1O	841 (1,545)	2.3 (7.6)	1.2 (47.5)	0.52	29,810 (12,816)	3.25	1.24	8.02	224 (1,778)	66 (524)	2.75	600	469 (1.09)	78.52	285	103 (0.24)	6,548 (15.23)	1,698 (3.95)
6-1P	838 (1,540)	2.3 (7.6)	1.2 (48.9)	0.52	29,810 (12,816)	3.25	1.24	8.02	223 (1,769)	60 (478)	2.53	606	477 (1.11)	78.00	285	107 (0.25)	6,582 (15.31)	1,707 (3.97)
6-2A	848 (1,559)	2.3 (7.6)	1.2 (48.4)	0.52	29,212 (12,559)	1.70	1.32	9.36	233 (1,852)	66 (521)	4.70	200	150 (0.35)	86.97	430	150 (0.35)	8,134 (18.92)	2,515 (5.85)
6-2B	848 (1,558)	2.3 (7.6)	1.3 (49.2)	0.57	29,212 (12,559)	1.70	1.32	9.36	234 (1,857)	66 (527)	4.74	204	155 (0.36)	86.63	430	155 (0.36)	8,113 (18.87)	2,506 (5.83)
6-2C	848 (1,559)	2.4 (7.7)	1.3 (49.3)	0.54	29,212 (12,559)	1.70	1.32	9.36	234 (1,858)	66 (520)	4.67	194	142 (0.33)	87.81	430	155 (0.36)	8,108 (18.86)	2,506 (5.83)
6-2D	848 (1,558)	2.2 (7.3)	1.2 (47.5)	0.55	29,170 (12,541)	2.53	1.31	8.82	202 (1,601)	51 (405)	2.99	301	259 (0.59)	85.47	350	138 (0.32)	8,392 (19.52)	2,313 (5.38)
6-2E	845 (1,553)	2.3 (7.4)	1.2 (47.8)	0.52	29,170 (12,561)	2.53	1.31	8.82	231 (1,834)	46 (366)	2.36	323	241 (0.56)	86.15	350	120 (0.28)	7,326 (17.04)	2,021 (4.70)
6-2F	844 (1,551)	2.3 (7.5)	1.2 (47.6)	0.52	29,170 (12,541)	2.53	1.31	8.82	238 (1,890	74 (589)	3.67	331	241 (0.56)	86.18	350	120 (0.28)	7,111 (16.54)	1,961 (4.56)
6-2G	843 (1,550)	2.2 (7.2)	1.2 (47.4)	0.55	29,170 (12,541)	2.53	1.31	8.82	230 (1,825)	54 (429)	2.77	363	267 (0.62)	84.73	350	120 (0.28)	7,365 (17.13)	2,029 (4.72)
6-2H	846 (1,555)	2.3 (7.4)	1.2 (48.2)	0.52	29,170 (12,541)	2.53	1.31	8.82	230 (1,823)	53 (420)	2.72	320	236 (0.55)	86.24	350	120 (0.28)	7,373 (17.15)	2,029 (4.72)
6-2I	846 (1,555)	2.2 (7.3)	1.2 (47.5)	0.55	29,815 (12,818)	2.27	1.32	8.14	234 (1,854)	50 (393)	2.73	398	284 (0.66)	81.45	320	107 (0.25)	7,833 (18.22)	2,498 (5.81)
6-2J	846 (1,555)	2.2 (7.2)	1.2 (47.2)	0.55	29,815 (12,818)	2.27	1.32	8.14	236 (1,873)	50 (393)	2.70	419	288 (0.67)	81.07	320	103 (0.24)	7,756 (18.04)	2,471 (5.75)
6-2K	847 (1,556)	2.3 (7.5)	1.2 (46.9)	0.52	29,815 (12,818)	2.27	1.32	8.14	235 (1,866)	44 (353)	2.44	379	275 (0.64)	81.96	320	107 (0.25)	7,702 (18.10)	2,481 (5.77)
6-2L	845 (1,552)	2.2 (7.2)	1.2 (47.2)	0.52	29,815 (12,818)	2.27	1.32	8.14	235 (1,863)	50 (393)	2.72	367	254 (0.59)	83.22	320	103 (0.24)	7,794 (18.13)	2,485 (5.78)
6-2M	847 (1,556)	2.4 (7.9)	1.2 (47.0)	0.50	29,536 (12,698)	2.58	1.34	9.35	248 (1,972)	57 (450)	2.59	514	370 (0.86)	78.79	290	99 (0.23)	8,396 (19.53)	3,246 (7.55)
6-2N	847 (1,556)	2.5 (8.2)	1.2 (47.6)	0.48	29,536 (12,698)	2.58	1.34	9.35	249 (1,974)	57 (456)	2.62	546	408 (0.95)	76.70	290	103 (0.24)	8,388 (19.51)	3,242 (7.54)

TABLE 81 (continued)

Test No.	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas residence time sec	Fuel characteristics					Sorbent characteristics		SO2			NOx		Particulate	
					Heating value k'/kg (Btu/lb)	% S	% N	% Ash	Feed rate g/s (lb/hr)	Feed rate g/s (lb/hr)	Ca/S ratio	ppm	ng/J (lb/10⁶ Btu)	SO2 retention	ppm	ng/J (lb/10⁶ Btu)	Cyclone inlet ng/J (lb/10⁶ Btu)	Cyclone outlet ng/J (lb/10⁶ Btu)
6-2O	844 (1,552)	2.5 (8.1)	1.2 (47.6)	0.48	29,536 (12,698)	2.58	1.34	9.35	248 (1,966)	57 (451)	2.61	509	378 (0.88)	78.29	290	103 (0.24)	8,422 (19.59)	3,255 (7.57)
6-2P	848 (1,559)	2.3 (7.6)	1.2 (47.5)	0.52	29,536 (12,698)	2.58	1.34	9.35	246 (1,954)	58 (463)	2.69	583	404 (0.94)	76.81	290	95 (0.22)	8,474 (19.71)	3,276 (7.62)
6-2Q	849 (1,560)	2.3 (7.4)	1.2 (48.4)	0.52	29,536 (12,698)	2.58	1.34	9.35	249 (1,977)	57 (450)	2.59	603	400 (0.93)	77.10	290	90 (0.21)	8,375 (19.48)	3,237 (7.53)
6-3A	848 (1,558)	2.9 (9.5)	1.2 (47.4)	0.41	29,743 (12,787)	2.87	1.24	8.50	248 (1,966)	56 (442)	2.28	551	470 (1.09)	75.72	370	146 (0.34)	9,162 (21.32)	3,727 (8.67)
6-3B	851 (1,564)	2.9 (9.4)	1.2 (48.2)	0.41	29,743 (12,787)	2.87	1.24	8.50	248 (1,967)	47 (455)	2.35	417	516 (1.20)	73.21	370	146 (0.34)	9,162 (21.31)	3,727 (8.67)
6-3C	850 (1,562)	2.8 (9.3)	1.3 (50.1)	0.46	29,743 (12,787)	2.87	1.24	8.50	246 (1,956)	55 (437)	2.27	498	417 (0.97)	78.35	370	146 (0.34)	9,213 (21.43)	3,749 (8.72)
6-3D	830 (1,562)	2.9 (9.6)	1.2 (48.5)	0.41	29,743 (12,787)	2.87	1.24	8.50	248 (1,969)	58 (461)	2.38	900	426 (0.99)	77.87	370	146 (0.34)	9,153 (21.29)	3,723 (8.66)
6-3E	852 (1,565	2.8 (9.2)	1.2 (48.9)	0.43	29,743 (12,787)	2.87	1.24	8.50	248 (1,967)	58 (462)	2.38	675	546 (1.27)	71.58	370	142 (0.33)	9,162 (21.31)	3,727 (8.67)
6-3F	853 (1,567	2.9 (9.4)	1.2 (48.1)	0.41	29,743 (12,787)	2.87	1.24	8.50	245 (1,944)	58 (461)	2.40	671	563 (1.31)	70.73	370	146 (0.34)	9,121 (21.56)	3,770 (8.77)
6-3G	845 (1,552)	2.8 (9.0)	1.2 (47.8)	0.43	29,770 (12,799)	2.18	1.23	8.04	253 (2,006)	60 (478)	3.03	282	226 (0.52)	84.75	375	138 (0.32)	9,377 (21.81)	3,616 (8.41)
6-3H	844 (1,551)	2.8 (9.1)	1.2 (47.8)	0.43	29,770 (12,799)	2.18	1.23	8.04	254 (2,014)	58 (464)	2.96	309	245 (0.57)	83.16	375	142 (0.33)	9,338 (21.72)	3,603 (8.38)
6-3I	844 (1,551)	2.7 (8.9)	1.2 (48.9)	0.44	29,770 (12,799)	2.18	1.23	8.04	252 (2,000)	59 (468)	3.01	299	236 (0.55)	83.92	375	138 (0.32)	9,407 (21.88)	3,629 (8.44)
6-3J	843 (1,550)	2.8 (9.0)	1.2 (47.4)	0.43	29,770 (12,799)	2.18	1.23	8.04	252 (2,003)	59 (470)	3.01	305	245 (0.57)	83.39	375	142 (0.33)	9,389 (21.84)	3,624 (8.43)

[*] By Beckman automatic infrared analyzer.

[†] Isokinetic sample drawn through fiberglass filter.

[‡] Although there was no limestone feed during these tests, residual sorbent in the bed probably accounts for the noted SO_2 removal efficiencies.

TABLE 82. EMISSION TEST DATA MEASURED DURING OPERATION OF B&W 3 FT × 3 FT FBC UNIT FIRING PITTSBURGH NO. 8 COAL[4]

Test no.	Date	Test duration (hr)	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas residence time (sec)	Fuel characteristics					Sorbent characteristics				Flue gas at WS inlet kg/h (lb/h)	Emissions characteristics						
							Heat value kJ/kg (Btu/lb)	% S	% Ash	% N	Feed rate kg/h (lb/h)	Type	Size μm (in. or M)	Feed rate kg/h (lb/h)	Ca/S ratio		SO₂ at WS inlet: ppm*	SO₂ at WS inlet: ng/J (lb/10⁶ Btu)	SO₂ at WS inlet: % reduction†	NOₓ at WS inlet‡: ppm	NOₓ at WS inlet‡: ng/J (lb/10⁶ Btu)	Particulate at WS inlet§: g/m (gr/acf)	Particulate at WS inlet§: ng/J (lb/10⁶ Btu)#
19	10/16 76	10.0	848 (1559)	2.56 (8.40)	0.47 (18.6)	0.18	29,275 (12,586)	3.06	9.32	0.86	227 (500)	Lowellville	6350 μm × 0 (1/4 in. × 0)	77 (170)	2.71	2,846 (6,274)	839	400 (0.93)	60.8	285	99 (0.23)	7.6 (3.3)	2,690 (6.27)
20	10/18/76	9.0	850 (1562)	2.57 (8.42)	0.50 (19.7)	0.20	29,275 (12,586)	3.06	9.32	0.86	227 (500)	Lowellville	6350 μm × 0 (1/4 in. × 0)	25 (56)	0.58	2,833 (6,246)	1,475	696 (1.62)	24.5	212	73 (0.17)	7.6 (3.3)	2,683 (6.24)
21	10/21/76	6.0	894 (1642)	2.56 (8.39)	0.43 (17.1)	0.17	29,275 (12,586)	3.06	9.32	0.86	193 (425)	Lowellville	6350 μm × 0 (1/4 in. × 0)	43 (96.5)	1.53	2,837 (6,254)	1,126	628 (1.46)	37.9	334	133 (0.31)	7.6 (3.3)	3,057 (7.11)
22	10/23/76	8.0	839 (1542)	2.57 (8.42)	0.51 (20.2)	0.20	29,275 (12,586)	3.06	9.32	0.86	209 (460)	Lowellville	6350 μm × 0 (1/4 in. × 0)	46 (102)	1.49	2,893 (6,377)	1,187	580 (1.35)	41.9	215	82 (0.19)	8.0 (3.5)	3,156 (7.34)
23	10/26/76	6.0	776 (1428)	2.45 (8.05)	0.59 (23.1)	0.24	29,375 (12,629)	2.86	9.43	0.86	222 (490)	Lowellville	6350 μm × 0 (1/4 in. × 0)	46 (102)	1.57	2,967 (6,542)	1,673	1,685 (3.92)	31.6	300	219 (0.51)	8.9 (3.9)	3,375 (7.85)
24	10/28/76	7.0	838 (1541)	3.63 (11.91)	0.62 (24.6)	0.17	29,375 (12,629)	2.86	9.43	0.86	344 (758)	Lowellville	6350 μm × 0 (1/4 in. × 0)	72 (159)	1.81	4,063 (8,958)	1,403	1,251 (2.91)	41.2	213	138 (0.32)	7.8 (3.4)	2,605 (6.06)
25	11/02/76	7.0	770 (1418)	1.40 (4.60)	0.87 (34.4)	0.42	29,375 (12,629)	2.86	9.43	0.86	112 (247)	Lowellville	6350 μm × 0 (1/4 in. × 0)	23 (51.5)	1.73	1,697 (3,741)	1,265	1,445 (3.36)	37.9	58	47 (0.11)	5.3 (2.3)	2,253 (5.24)
26	11/09/76	10.0	829 (1525)	2.49 (8.16)	0.42 (16.5)	0.17	29,484 (12,676)	2.86	9.43	0.76	211 (464)	Lowellville	2380 μm × 0 (8 mesh × 0)	44 (96.4)	2.16	2,844 (6,269)	879	890 (2.07)	58.7	213	155 (0.36)	8.9 (3.9)	3,315 (7.71)
27	11/10/76	8.0	842 (1548)	2.52 (8.28)	0.42 (16.5)	0.17	29,484 (12,676)	2.86	9.43	0.76	209 (460)	Lowellville	2380 μm × 0 (8 mesh × 0)	43 (95)	2.28	2,868 (6,322)	849	877 (2.04)	60.8	285	211 (0.49)	8.9 (3.9)	3,461 (8.05)
28	11/11/76	7.5	858 (1576)	2.58 (8.45)	0.42 (16.5)	0.16	29,484 (12,676)	2.86	9.43	0.76	218 (480)	Lowellville	2380 μm × 0 (8 mesh × 0)	68 (150)	3.51	2,892 (6,375)	589	585 (1.36)	76.6	322	228 (0.53)	9.8 (4.3)	3,689 (8.58)
29	11/12/76	8.0	845 (1553)	2.54 (8.34)	0.43 (17.1)	0.17	29,484 (12,676)	2.86	9.43	0.76	200 (440)	Lowellville	2380 μm × 0 (8 mesh × 0)	21 (47)	1.11	2,876 (6,341)	1,163	1,238 (2.88)	39.2	233	181 (0.42)	7.8 (3.4)	3,166 (7.36)
30	11/13/76	5.5	850 (1562)	2.54 (8.34)	0.43 (16.9)	0.17	29,484 (12,676)	2.86	9.43	0.76	204 (450)	Lowellville	2380 μm × 0 (8 mesh × 0)	21 (47)	1.11	2,861 (6,308)	1,502	1,582 (3.68)	33.3	303	228 (0.53)	7.3 (3.2)	2,898 (6.74)
31	11/16/76	8.5	819 (1506)	2.48 (8.13)	0.40 (15.8)	0.16	29,484 (12,676)	2.86	9.43	0.76	222 (490)	Lowellville	1000 μm × 0 (16 mesh × 0)	44 (98)	2.46	2,869 (6,325)	1,035	1,006 (2.34)	58.1	238	168 (0.39)	10.5 (4.6)	4,086 (9.50)
32	11/17/76	7.5	826 (1518)	2.48 (8.15)	0.42 (16.4)	0.17	29,484 (12,676)	2.86	9.43	0.76	218 (480)	Lowellville	1000 μm × 0 (16 mesh × 0)	66 (145)	3.49	2,852 (6,287)	676	666 (1.55)	74.0	0	0 (0)	10.1 (4.4)	3,723 (8.66)
33	11/19/76	6.5	845 (1553)	2.51 (8.22)	0.42 (16.4)	0.17	29,484 (12,676)	2.86	9.43	0.76	209 (460)	Lowellville	1000 μm × 0 (16 mesh × 0)	20 (45)	0.87	2,840 (6,262)	1,546	1,582 (3.68)	29.1	263	193 (0.45)	8.0 (3.5)	3,078 (7.16)
35	12/08/76	8.5	845 (1553)	2.59 (8.50)	0.41 (16.2)	0.16	29,484 (12,676)	2.86	9.43	0.76	209 (460)	Lowellville	Pulverized	64 (140)	2.05	2,908 (6,412)	992	1,040 (2.42)	45.9	246	185 (0.43)	14.6 (6.4)	5,765 (13.41)
36	12/09/76	4.5	819 (1507)	1.55 (5.08)	0.29 (11.4)	0.19	29,484 (12,676)	2.86	9.43	0.76	154 (340)	Lowellville	Pulverized	45 (100)	2.25	1,802 (3,972)	930	816 (1.90)	61.0	296	185 (0.43)	16.5 (7.2)	5,434 (12.64)
37	12/10/76	4.5	851 (1563)	2.57 (8.44)	0.34 (13.3)	0.13	29,484 (12,676)	2.86	9.43	0.76	245 (540)	Lowellville	Pulverized	49 (107)	2.38	2,876 (6,341)	895	791 (1.84)	59.5	198	125 (0.29)	16.7 (7.3)	5,537 (12.88)

TABLE 82 (continued)

Test no.	Date	Test duration (hr)	Bed temperature °C (°F)	Superficial gas velocity m/s (f/s)	Bed depth m (in.)	Gas residence time (sec)	Fuel characteristics: Heat value kJ/kg (Btu/lb)	Fuel characteristics: % S	Fuel characteristics: % Ash	Fuel characteristics: % N	Fuel characteristics: Feed rate kg/h (lb/h)	Sorbent characteristics: Type	Sorbent characteristics: Size μm (in. or M)	Sorbent characteristics: Feed rate kg/h (lb/h)	Sorbent characteristics: Ca/S ratio	Flue gas at WS inlet kg/h (lb/h)	Emissions characteristics: SO_2 at WS inlet: ppm*	Emissions characteristics: SO_2 at WS inlet: ng/J (lb/10^6 Btu)	Emissions characteristics: SO_2 at WS inlet: % reduction†	Emissions characteristics: NO_x at WS inlet‡: ppm	Emissions characteristics: NO_x at WS inlet‡: ng/J (lb/10^6 Btu)	Emissions characteristics: Particulate at WS inlet§: g/m (gr/scf)	Emissions characteristics: Particulate at WS inlet§: ng/J (lb/10^6 Btu)#
38	12/13/76	7.0	865 (1589)	2.64 (8.67)	0.43 (17.1)	0.16	29,484 (12,676)	2.86	9.43	0.76	240 (530)	$Ca(OH)_2$	44 μm = 0 (325 mesh = 0)	49 (107)	2.18	2,929 (6,458)	387	355 (0.82)	78.2	199	129 (0.30)	21.0 (9.2)	6,453 (15.01)
40	12/16/76	9.5	829 (1525)	1.49 (4.90)	0.31 (12.2)	0.21	29,484 (12,676)	2.86	9.43	0.76	134 (295)	$Ca(OH)_2$	44 μm = 0 (325 mesh = 0)	19 (41)	1.88	1,721 (3,793)	536	507 (1.18)	73.2	313	219 (0.51)	19.7 (8.6)	7,145 (16.62)
41	12/17/76	12.5	845 (1553)	2.56 (8.41)	0.37 (14.5)	0.14	29,484 (12,676)	2.86	9.43	0.76	227 (500)	$Ca(OH)_2$	44 μm = 0 (325 mesh = 0)	31 (69)	1.69	2,903 (6,401)	484	466 (1.08)	73.6	265	185 (0.43)	18.5 (8.1)	6,698 (15.58)
42	01/11/77	8.0	842 (1547)	2.66 (8.74)	0.45 (17.6)	0.17	29,484 (12,676)	2.86	9.43	0.76	234 (515)	Lowellville	2380 μm = 0 (8 mesh = 0)	79 (175)	3.36	3,015 (6,646)	758	735 (1.71)	60.3	318	224 (0.52)	10.1 (4.4)	3,667 (8.53)
43	01/12/77	9.0	841 (1546)	2.64 (8.67)	0.44 (17.2)	0.17	29,484 (12,676)	2.86	9.43	0.76	234 (515)	Lowellville	1000 μm = 0 (16 mesh = 0)	79 (175)	3.33	3,015 (6,646)	858	834 (1.94)	66.1	340	236 (0.55)	11.6 (5.0)	4,170 (9.70)
44	01/13/77	5.5	841 (1546)	2.63 (8.64)	0.44 (17.5)	0.17	29,115 (12,517)	3.12	9.74	1.23	244 (539)	Lowellville	Pulverized	92 (202)	2.76	3,014 (6,645)	879	825 (1.92)	56.9	224	150 (0.35)	22.2 (9.7)	7,825 (18.20)
45	01/13/77	1.5	861 (1582)	2.62 (8.61)	0.46 (18.3)	0.18	29,115 (12,517)	3.12	9.74	1.23	222 (490)	$Ca(OH)_2$	44 μm = 0 (325 mesh = 0)	18 (40)	0.99	2,968 (6,529)	1,521	1,543 (3.59)	34.1	257	189 (0.44)	13.0 (5.7)	4,970 (11.56)
46	01/20/77	7.0	837 (1538)	2.59 (8.49)	0.45 (17.9)	0.18	29,115 (12,517)	3.12	9.74	1.23	224 (494)	Greer	2380 μm = 0 (8 mesh = 0)	127 (279)	3.62	2,977 (6,563)	401	404 (0.94)	81.7	360	262 (0.61)	9.8 (4.3)	3,736 (8.69)
47	01/21/77	6.0	838 (1540)	2.60 (8.54)	0.43 (16.8)	0.16	29,115 (12,517)	3.12	9.74	1.23	220 (485)	Greer	1000 μm = 0 (16 mesh = 0)	116 (256)	3.94	2,982 (6,574)	380	391 (0.91)	85.0	296	219 (0.51)	9.4 (4.1)	3,637 (8.46)
48	01/22/77	6.0	843 (1549)	2.62 (8.61)	0.37 (14.7)	0.14	29,115 (12,517)	3.12	9.74	1.23	227 (500)	Greer	Pulverized	116 (255)	2.70	2,995 (6,603)	1,140	1,148 (2.67)	48.3	246	176 (0.41)	28.1 (12.3)	10,623 (24.71)
49	01/26/77	6.0	840 (1544)	2.58 (8.46)	0.43 (17.1)	0.17	29,115 (12,517)	3.12	9.74	1.23	225 (496)	Grove	2380 μm = 0 (8 mesh = 0)	104 (230)	4.65	2,937 (6,475)	536	533 (1.24)	77.4	307	219 (0.51)	9.2 (4.0)	3,500 (8.14)
50	01/27/77	9.0	846 (1554)	2.62 (8.60)	0.42 (16.5)	0.16	29,115 (12,517)	3.12	9.74	1.23	230 (507)	Grove	1000 μm = 0 (16 mesh = 0)	113 (250)	4.93	2,964 (6,535)	467	460 (1.07)	81.1	309	219 (0.51)	9.4 (4.1)	3,457 (8.04)
51	02/07/77	6.0	837 (1539)	2.64 (8.66)	0.36 (14.0)	0.13	29,115 (12,517)	3.12	9.74	1.23	222 (490)	Grove	Pulverized	113 (250)	3.95	3,005 (6,625)	765	765 (1.78)	59.1	206	150 (0.35)	39.4 (17.2)	15,215 (35.39)

*By DuPont Model 411, UV light absorption.

†Short gas residence time (0.20 sec and less), low freeboard and absence of primary recycle limited SO_2 removal efficiencies.

‡By Teco Model 10A chemiluminescence monitor.

§Using five-point traverse and glass filter.

#Low freeboard and absence of primary recycle accounts for high particle loadings at WS inlet.

TABLE 83. EMISSION SOURCE TEST DATA: NCB-CRE 3 FT × 1.5 FT ATMOSPHERIC FBC[5,6]

					Fuel characteristics			Sorbent characteristics			Emissions characteristics				
Test no.	Bed temperature °C (°F)	Gas velocity m/s (ft/s)	Bed depth m (ft)	Gas phase residence time sec	Heat value* kJ/kg (Btu/lb)	% S	% Ash	Type	Size† median μm	Ca/S	SO_2‡ ppm	SO_2 ng/J (lb/10^6 Btu)§	% control	NO_x# ppm	ng/J (lb/10^6 Btu)§
1.1	799 (1470)	1.2 (4.0)	0.70 (2.3)	0.58	35,062 (15,074)	2.8	13.5	-	-	0	1,750	1,596 (3.7)	0	NR	NR (NR)
1.2	799 (1470)	1.2 (4.0)	0.70 (2.3)	0.58	35,062 (15,074)	2.8	13.5	-	-	0	2,050	1,596 (3.7)	0	NR	NR (NR)
Datum	849 (1560)	1.2 (4.0)	0.67 (2.2)	0.58	35,062 (15,074)	2.8	13.5	-	-	0	2,100	1,596 (3.7)	0	NR	NR (NR)
Datum	799 (1470)	0.91 (3.0)	0.70 (2.3)	0.77	35,062 (15,074)	2.8	13.5	-	-	0	2,020	1,596 (3.7)	0	NR	NR (NR)
1.3	799 (1470)	1.2 (4.0)	0.67 (2.2)	0.58	35,062 (15,074)	2.8	13.5	Limestone 18	210	2.2	400	301 (0.70)	81	NR	NR (NR)
1.4	749 (1380)	1.2 (4.0)	0.67 (2.2)	0.58	35,062 (15,074)	2.8	13.5	Limestone 18	210	2.2	1,020	796 (1.9)	50	NR	NR (NR)
1.5	849 (1560)	1.2 (4.0)	0.67 (2.2)	0.58	35,062 (15,074)	2.8	13.5	Limestone 18	210	2.2	360	271 (0.63)	83	NR	NR (NR)
1.6	849 (1560)	1.2 (4.0)	0.67 (2.2)	0.58	35,062 (15,074)	2.8	13.5	Limestone 18	210	1.3	880	671 (1.6)	58	NR	NR (NR)
1.7	849 (1560)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	Limestone 18	210	2.2	510	353 (0.82)	76	NR	NR (NR)
1.8	849 (1560)	1.2 (4.0)	0.67 (2.2)	0.58	35,062 (15,074)	2.8	13.5	Limestone 18	210	3.3	180	142 (0.33)	91	NR	NR (NR)
1.9	849 (1560)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	Limestone 18	210	1.2	840	637 (1.5)	60	NR	NR (NR)
1.10	799 (1470)	0.91 (3.0)	0.70 (2.3)	0.77	35,062 (15,074)	2.8	13.5	Limestone 18	210	2.2	330	254 (0.59)	84	NR	NR (NR)
1.11	799 (1470)	0.94 (3.1)	0.64 (2.1)	0.68	35,062 (15,074)	2.8	13.5	Limestone 18	210	3.3	42	30 (0.07)	98	NR	NR (NR)
3.1	849 (1560)	2.4 (7.9)	0.64 (2.1)	0.27	33,437 (14,375)	1.3	18.2	-	-	0	1,480	777 (1.8)	0	NR	NR (NR)
3.2	849 (1560)	2.4 (8.0)	0.64 (2.1)	0.26	33,437 (14,375)	1.3	18.2	U.K. Limestone	300 - 400	1.8	910	482 (1.1)	38	NR	NR (NR)
3.3	849 (1560)	2.4 (8.0)	0.64 (2.1)	0.26	33,437 (14,375)	1.3	18.2	U.K. Limestone	300 - 400	2.1	740	389 (0.90)	49	NR	NR (NR)
3.4	849 (1560)	2.4 (7.9)	0.67 (2.2)	0.28	33,437 (14,375)	1.3	18.2	U.K. Limestone	300 - 400	2.8	540	280 (0.65)	64	NR	NR (NR)
3.5	849 (1560)	2.4 (8.0)	1.16 (3.8)	0.48	33,437 (14,375)	1.3	18.2	U.K. Limestone	300 - 400	2.8	540	280 (0.65)	64	NR	NR (NR)
3.6	849 (1560)	1.2 (4.1)	0.7 (2.3)	0.58	33,437 (14,375)	1.3	18.2	U.K. Limestone	300 - 400	3.0	420	218 (0.51)	72	NR	NR (NR)
Datum	849 (1560)	2.5 (8.1)	0.64 (2.1)	0.26	35,062 (15,074)	2.8	13.5	-	-	0	1,830	1,596 (3.7)	0	424	266 (0.62)

TABLE 83 (continued)

Test no.	Bed temperature °C (°F)	Gas velocity m/s (ft/s)	Bed depth m (ft)	Gas phase residence time sec	Fuel characteristics			Sorbent characteristics			Emissions characteristics				
					Heat value* kJ/kg (Btu/lb)	% S	% Ash	Type	Size† median µm	Ca/S	SO_2† ppm	SO_2 ng/J (lb/10^6 Btu)§	% control	NO_x# ppm	ng/J (lb/10^6 Btu)§
Datum	849 (1560)	1.2 (4.0)	0.64 (2.1)	0.53	36,062 (15,074)	2.8	13.5	-	-	0	2,100	1,596 (3.7)	0	NR	NR (NR)
Datum	849 (1560)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	-	-	0	2,400	1,596 (3.7)	0	305	146 (0.35)
2.1	849 (1560)	2.5 (8.1)	0.64 (2.1)	0.26	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	1.1	1,200	1,054 (2.5)	34	470	297 (0.69)
2.2	849 (1560)	2.5 (8.1)	0.64 (2.1)	0.26	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	2.3	900	703 (1.6)	56	515	289 (0.67)
2.3	849 (1560)	2.4 (8.0)	0.64 (2.1)	0.26	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	2.9	650	511 (1.2)	68	NR	NR (NR)
5.1	849 (1560)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	1.7	850	559 (1.3)	65	NR	NR (NR)
5.2	849 (1560)	2.4 (8.0)	0.67 (2.2)	0.28	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	1.6	1,180	783 (1.8)	51	445	212 (0.49)
5.3	849 (1560)	2.4 (7.8)	1.07 (3.5)	0.45	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	1.9	1,040	687 (1.6)	57	NR	NR (NR)
5.4	849 (1560)	2.4 (8.0)	1.10 (3.6)	0.45	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	5.7	80	48 (0.11)	97	560	242 (0.56)
5.5	849 (1560)	2.4 (8.0)	1.10 (3.6)	0.45	35,062 (15,074)	2.8	13.5	Limestone 18	350 - 450	6.0	11	9 (0.02)	100	550, 325	323, 191 (0.75), (0.44)
2.4	849 (1560)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	Limestone 18	-125	1.0	1,050	814 (1.9)	49	NR	NR (NR)
2.5	849 (1560)	0.91 (3.0)	0.61 (2.0)	0.67	35,062 (15,074)	2.8	13.5	Limestone 18	-125	1.0	1,050	814 (1.9)	49	NR	NR (NR)
Datum	799 (1470)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	-	-	0	2,240	1,596 (3.7)	0	NR	NR (NR)
Datum	849 (1560)	2.4 (8.0)	0.82 (2.7)	0.34	35,062 (15,074)	2.8	13.5	-	-	0	2,190	1,596 (3.7)	0	390	204 (0.47)
4.1	799 (1470)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	Dolomite 1337	100 - 125	3.1	380	271 (0.63)	83	234	120 (0.28)
4.2	749 (1380)	1.2 (4.1)	0.64 (2.1)	0.51	35,062 (15,074)	2.8	13.5	Dolomite 1337	100 - 125	2.6	620	447 (1.0)	72	NR	NR (NR)
4.3	849 (1560)	1.2 (3.8)	0.64 (2.1)	0.55	35,062 (15,074)	2.8	13.5	Dolomite 1337	100 - 125	2.7	600	431 (1.0)	73	244	126 (0.29)
4.4	799 (1470)	1.2 (4.0)	0.64 (2.1)	0.53	35,062 (15,074)	2.8	13.5	Dolomite 1337	100 - 125	2.7	580	415 (0.97)	74	NR	NR (NR)
4.5	799 (1470)	0.64 (2.1)	1.13 (3.7)	1.76	35,062 (15,074)	2.8	13.5	Dolomite 1337	100 - 125	2.2	380	271 (0.63)	83	360	185 (0.43)

TABLE 83 (continued)

Test no.	Bed temperature °C (°F)	Gas velocity m/s (ft/s)	Bed depth m (ft)	Gas phase residence time sec	Fuel characteristics			Sorbent characteristics			Emissions characteristics				
					Heat value* kJ/kg (Btu/lb)	% S	% Ash	Type	Size† median μm	Ca/S	SO_2‡ ppm	SO_2 ng/J (lb/10^6 Btu)§	% control	NO_x# ppm	ng/J (lb/10^6 Btu)§
4.6	799 (1470)	0.67 (2.2)	1.16 (3.8)	1.73	35,062 (15,074)	2.8	13.5	Dolomite 1337	100 - 125	1.6**	20	16 (0.04)	99	392	225 (0.52)
6.1	849 (1560)	2.4 (8.0)	0.82 (2.7)	0.34	35,062 (15,074)	2.8	13.5	Dolomite 1337	875 - 1000	2.5	840	575 (1.3)	64	NR	NR (NR)
6.2	849 (1560)	2.4 (8.0)	0.82 (2.7)	0.34	35,062 (15,074)	2.8	13.5	Dolomite 1337	875 - 1000	5.4	280	208 (0.48)	87	390	208 (0.48)
6.3	849 (1560)	2.5 (8.1)	1.22 (4.0)	0.49	35,062 (15,074)	2.8	13.5	Dolomite 1337	875 - 1000	5.3	260	192 (0.45)	88	360	191 (0.44)
6.4	849 (1560)	2.4 (8.0)	2.13 (7.0)	0.88	35,062 (15,074)	2.8	13.5	Dolomite 1337	875 - 1000	5.2	155	112 (0.26)	93	400	207 (0.48)
6.5	849 (1560)	2.4 (8.0)	1.68 (5.5)	0.69	35,062 (15,074)	2.8	13.5	Dolomite 1337	875 - 1000	5.0	280	208 (48)	87	425	226 (0.53)

*Feed rate varied between 75 and 300 lb/h; heating value measured on dry ash free basis.

†Size range for all limestone is -1680 μm × 0

‡By continuous online Hartman-Braun infrared analyzer, the iodine method,[10] and the hydrogen peroxide method.[11]

§Estimated by GCA.

#By a modified Saltzman's method,[12] and the BCURA NO_x box.[13]

**With fines recycle.

NR = Not reported.

TABLE 84. PER-FBM EMISSION SOURCE TEST DATA RECORDED IN TESTS CONDUCTED FROM LATE 1967 THROUGH 1969[7]

								Coal					Sorbent				Flue gas						
Test number	Test condition number	Static bed depth cm (in.)	Estimated[a] expanded bed depth cm (in.)	Air feed rate kg/hr (lb/hr)	Calculated[b] gas velocity m/s (ft/s)	Estimated[a] gas residence time (sec)	Bed temperature °C (°F)	Type	Feed rate kg/hr (lb/hr)	% S	% Ash	HHV kJ/kg (Btu/lb)	Type	Size μm	Feed rate kg/hr (lb/hr)	Ca/S ratio	SO_2 IBA[c] ppm	SO_2 wet chemical analysis[d] ppm	SO_2 ng/J[e] (lb/10⁶ Btu)	SO_2 reduction[f] %	NO IBA[c] ppm	NO_x PDSB ppm	NO_x ng/J[h] (lb/10⁶ Btu)
1[i]	1	50.8 (20)	76 (30)	3,273 (7,200)	3.1 (10.2)	0.25	838 (540)	East Kentucky	364 (800)	1.0	8.0		-				680			0	200		
	2	50.8 (20)	76 (30)	3,273 (7,200)	3.1 (10.2)	0.25	838 (540)		364 (800)	1.0	8.0		-				600			0	220		
	3	50.8 (20)	76 (30)	3,273 (7,200)	3.1 (10.2)	0.25	838 (540)		364 (800)	1.0	8.0		-				500			0	260		
	4	50.8 (20)	76 (30)	3,273 (7,200)	3.1 (10.2)	0.25	838 (540)		364 (800)	1.0	8.0		-				620			0	240		
2[j]	1	30.5 (12)	46 (18)	3,364 (7,400)	3.6 (11.9)	0.13	982 (1,800)	Ohio #8 Seam Unwashed	377 (830)	4.5	10.7	30,084 (12,934)	-				700		3,009 (7.0)	0	250		158 (0.37)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.4 (11.1)	0.23	899 (1,650)		366 (800)	4.5	10.7	30,084 (12,934)	-				3,400		3,009 (7.0)	0	300		190 (0.44)
	3	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)		366 (800)	4.5	10.7	30,084 (12,934)	1337B[h,i]	-2,830 +1,410	168 (370)	1.75	1,500		1,369 (3.2)	54.5	300		190 (0.44)
	4	50.8 (20)	76 (30)	3,364 (7,400)	3.1 (10.3)	0.26	816 (1,500)		364 (800)	4.5	10.7	30,084 (12,934)	-				800		3,009 (7.0)	0	340		215 (0.5)
3	1[g]	50.8 (20)	76 (30)	3,455 (7,600)	3.4 (11.2)	0.22	882 (1,620)	Ohio # 8 Seam Unwashed	377 (830)	4.5	10.7	30,084 (12,934)	-				3,600		3,009 (7.0)	0	380		228 (0.53)
	2	50.8 (20)	76 (30)	3,455 (7,600)	3.3 (10.8)	0.23	838 (1,540)		395 (870)	4.5	10.7	30,084 (12,934)	1337B	-2,830 +1,410	145 (320)	1.6	2,050		1,715 (4.0)	43.0	360		216 (0.5)
4	1	38.1 (15)	58 (23)	3,455 (7,600)	3.7 (12.0)	0.21	960 (1,760)	Ohio #8 Seam Unwashed	377 (830)	4.5	10.7	30,084 (12,934)	-				3,800	3,340	3,009 (7.0)	0	360	134	205 (0.48)
	2	38.1 (15)	58 (23)	3,455 (7,600)	3.6 (11.9)	0.16	949 (1,740)					30,084 (12,934)	1337B	-2,830 +1,410	115 (254)	1.15	3,700		2,931 (6.8)	2.6	340		194 (0.45)
	3	38.1 (15)	58 (23)	3,455 (7,600)	3.6 (11.8)	0.16	938 (1,720)		377 (830)	4.5	10.7	30,084 (12,934)			161 (355)	1.60	3,300		2,612 (6.1)	13.2	340		194 (0.45)
	4[g]	38.1 (15)	58 (23)	3,455 (7,600)	3.6 (11.8)	0.16	938 (1,720)		377 (830)	4.5	10.7	30,084 (12,934)			218 (480)	2.20	2,700		2,136 (5.0)	29.0	300		170 (0.40)

TABLE 84 (continued)

								Coal					Sorbent				Flue gas						
Test number	Test condition number	Static bed depth cm (in.)	Estimated^a expanded bed depth cm (in.)	Air feed rate kg/hr (lb/hr)	Calculated^b gas velocity m/s (ft/s)	Estimated^a gas residence time (sec)	Bed temperature °C (°F)	Type	Feed rate kg/hr (lb/hr)	% S	% Ash	HHV kJ/kg (Btu/lb)	Type	Size µm	Feed rate kg/hr (lb/hr)	Ca/S ratio	SO_2 IRA^c ppm	SO_2 wet chemical analysis^d ppm	SO_2 ng/J^e (lb/10^6 Btu)	SO_2 reduction^f %	NO IRA^c ppm	NO_x PDS ppm	NO_x ng/J^h (lb/10^6 Btu)
5	1	50.8 (20)	76 (30)	3,091 (6,800)	3.0 (10.0)	0.25	871 (1,600)	Ohio #8 Seam Unwashed	327 (720)	4.5	10.7	30,084 (12,934)	-				3,800	3,500	3,009 (7.0)	0	280	162	159 (0.37)
	2	50.8 (20)	76 (30)	3,091 (6,800)	3.0 (9.8)	0.26	854 (1,570)		327 (720)	4.5	10.7	30,084 (12,934)	1337B	-2,830 +1,410	100 (220)	1.13	3,000		2,365 (5.5)	21.0	280		159 (0.37)
	3	50.8 (20)	76 (30)	3,091 (6,800)	3.0 (9.8)	0.26	854 (1,570)		327 (720)	4.5	10.7	30,084 (12,934)			144 (317)	1.65	2,400		1,892 (4.4)	37.0	280		159 (0.37)
	4	50.8 (20)	76 (30)	3,091 (6,800)	3.0 (9.8)	0.26	854 (1,570)		327 (720)	4.5	10.7	30,084 (12,934)			209 (460)	2.40	2,000	1,850	1,548 (3.6)	48.0	280		159 (0.37)
7	1	50.8 (20)	76 (30)	3,409 (7,500)	3.4 (11.1)	0.23	882 (1,620)	Ohio #8 Seam Unwashed	355 (780)	4.5	10.7	30,084 (12,934)	-				3,900	3,640	3,009 (7.0)	0	280	189	159 (0.37)
	2	50.8 (20)	76 (30)		3.1 (10.3)	0.24	806 (1,480)		355 (780)	4.5	10.7	30,084 (12,934)	[illegible]^e	-2,830 +1,410	80 (175)	1.11	2,800		2,150 (5.0)	28.2	280		143 (0.33)
	3	50.8 (20)	76 (30)		3.1 (10.3)	0.24	804 (1,480)		355 (780)	4.5	10.7	30,084 (12,934)			120 (265)	1.70	2,150		1,634 (3.8)	45.0	220		120 (0.28)
	4^p	50.8 (20)	76 (30)		3.1 (10.3)	0.24	804 (1,480)		355 (780)	4.5	10.7	30,084 (12,934)			80 (175)	1.11	2,400	2,150	1,806 (4.2)	39.0	160		87 (0.20)
17	1	50.8 (20)	76 (30)	3,565 (7,800)	3.5 (11.6)	0.22	871 (1,600)	Ohio #8 Seam Unwashed	341 (750)	4.5	10.7	30,084 (12,934)	-				3,900	3,870	3,009 (7.0)	0	280		159 (0.37)
	2	50.8 (20)	76 (30)	3,565 (7,800)	3.4 (11.3)	0.22	860 (1,580)		373 (820)	4.5	10.7	30,084 (12,934)	1359m^g	-44	29 (63)	0.72	2,800	2,690	2,150 (5.0)	28.2	240	285	132 (0.31)
	3	50.8 (20)	76 (30)	3,565 (7,800)	3.4 (11.3)	0.22	860 (1,580)		364 (800)	4.5	10.7	30,084 (12,934)			38 (84)	0.98	2,300	2,180	1,462 (3.4)	41.0	200		91 (0.21)
	4^p	50.8 (20)	76 (30)	3,565 (7,800)	3.3 (10.8)	0.23	810 (1,490)		382 (840)	4.5	10.7	30,084 (12,934)			38 (84)	0.93	1,800	1,820	1,376 (3.2)	54.0	200		110 (0.26)
18^i	1	33.0 (13)	51 (20)	3,565 (7,800)	3.8 (12.9)	0.13	1,021 (1,870)	Ohio #8 Seam Unwashed	377 (830)	2.6	7.2	31,820 (13,680)	-				2,800		1,634 (3.8)	0	280		117 (0.27)
	2	33.0 (13)	51 (20)	3,565 (7,800)	3.9 (12.7)	0.13	1,004 (1,840)			2.6	7.2	31,820 (13,680)	-				2,650		1,634 (3.8)	0	300		133 (0.31)
	3	33.0 (13)	51 (20)	3,565 (7,800)	3.8 (12.5)	0.13	982 (1,800)			2.6	7.2	31,820 (13,680)	-				2,500		1,634 (3.8)	0	320	323	150 (0.35)
	4	33.0 (13)	51 (20)	3,565 (7,800)	3.8 (12.4)	0.13	966 (1,770)		364 (800)	2.6	7.2	31,820	-				2,300		1,634 (3.8)	0	340		174 (0.40)

TABLE 84 (continued)

Test number	Test condition number	Static bed depth cm (in.)	Estimated[a] expanded bed depth cm (in.)	Air feed rate kg/hr (lb/hr)	Calculated[b] gas velocity m/s (ft/s)	Estimated[a] gas residence time (sec)	Bed temperature °C (°F)	Coal					Sorbent				Flue gas							Fly ash[i] kg/hr (lb/hr)	Fly ash[i] ng/J[h] (lb/10 Btu)
								Type	Feed rate kg/hr (lb/hr)	% S	% Ash	HHV kJ/kg (Btu/lb)	Type	Size µm	Feed rate kg/hr (lb/hr)	Ca/S ratio	SO_2 IRA[c] ppm	SO_2 wet chemical analysis[d] ppm	SO_2 ng/J[e] (lb/10^6 Btu)	SO_2 reduction[f] %	NO IRA[c] ppm	NO_x PDS[g] ppm	NO_x ng/J[h] (lb/10^6 Btu)		
20	1	33.0 (13)	49.5 (19.5)	3,545 (7,800)	3.8 (12.4)	0.13	966 (1,770)	Ohio #8 Seam Washed	364 (800)	2.6	7.2	31,820 (13,680)	-			0	2,200	2,130	1,634 (3.8)	0	340		174 (0.40)	NR	
	2	33.0 (13)	49.5 (19.5)	3,545 (7,800)	3.8 (12.6)	0.13	988 (1,810)		400 (880)	2.6	7.2	31,820 (13,680)	1337H	-44	51 (112)	1.17	1,650	1,590	1,226 (2.9)	25.0	260	280	139 (0.32)		
	3	33.0 (13)	49.5 (19.5)	3,545 (7,800)	3.8 (12.4)	0.13	966 (1,770)		409 (900)	2.6	7.2	31,820 (13,680)			65 (144)	1.46	1,300		980 (2.3)	40.0	230		125 (0.29)		
	4[p]	33.0 (13)	49.5 (19.5)	3,545 (7,800)	3.7 (12.0)	0.14	932 (1,710)		409 (900)	2.6	7.2	31,820 (13,680)			65 (144)	1.45	900		670 (1.6)	59.0	230		125 (0.29)		
21	1	48.3 (19)	72.4 (28.5)	3,545 (7,800)	3.6 (11.9)	0.20	916 (1,680)	Ohio #8 Seam Washed	400 (880)	2.6	7.2	31,820 (13,680)	-			0	2,250	2,200	1,634 (3.8)	0	280	303	146 (0.34)	5.5 (12.2)	434 (1.01)
	2	48.3 (19)	72.4 (28.5)	3,545 (7,800)	3.6 (11.9)	0.20	916 (1,680)		400 (880)	2.6	7.2	31,820 (13,680)	1337H	-44	60 (132)	1.37	2,250	850	719 (1.7)	56.0	280		146 (0.34)	6.3 (13.8)	494 (1.15)
	3[p]	48.3 (19)	72.4 (28.5)	3,545 (7,800)	3.5 (11.4)	0.21	871 (1,600)		400 (800)	2.6	7.2	31,820 (13,680)			60 (132)	1.37	2,250		621 (1.4)	62.0					
22	1	50.8 (20)	76 (30)	3,545 (7,800)	4.1 (13.4)	0.19	899 (1,950)	Ohio #8 Seam Washed	420 (925)	2.6	7.2	31,820 (13,680)				0	2,300	2,250	1,634 (3.8)	0	280	275	143 (0.33)	5.3 (11.6)	396 (0.92)
	2	50.8 (20)	76 (30)	3,545 (7,800)	4.1 (13.4)	0.19	899 (1,950)		416 (915)	2.6	7.2	31,820 (13,680)	1337H	-44	66 (145)	1.46	750	680	523 (1.2)	68.0	280		143 (0.33)	6.5 (14.2)	486 (1.13)
23	1	48.3 (19)	72.4 (28.5)	3,364 (7,400)	3.4 (11.0)	0.22	888 (1,630)	Ohio #8 Seam Washed	364 (800)	2.6	7.2	31,820 (13,680)				0	2,300		1,634 (38)	0	300	280	153 (0.36)	6.3 (13.9)	546 (1.27)
	2	48.3 (19)	72.4 (28.5)	3,364 (7,400)	3.2 (10.6)	0.22	843 (1,550)		364 (800)	2.6	7.2	31,820 (13,680)	1337B	-44	128 (282)	2.6	800		567 (1.3)	65.3	300		153 (0.36)	7.0 (15.3)	602 (1.40)
24	1	55.9 (22)	83.8 (33)	3,364 (7,400)	3.3 (10.8)	0.25	871 (1,600)	Ohio #8 Seam Washed	345 (760)	2.6	7.2	31,820 (13,680)				0	3,400	2,400	1,634 (3.8)	0	260	285	127 (0.30)	4.2 (9.3)	383 (0.89)
	2	55.9 (22)	83.8 (33)	3,364 (7,400)	3.3 (10.8)	0.25	871 (1,600)		364 (800)	2.6	7.2	31,820 (13,680)	1337B	-44	127 (280)	2.4	650	560	297 (0.7)	81.8	260		127 (0.30)	6.2 (13.6)	533 (1.24)
	3	55.9 (22)	83.8 (33)	3,364 (7,400)	3.3 (10.8)	0.25	871 (1,600)		364 (800)	2.6	7.2	31,820 (13,680)			118 (260)	2.2	500	620	369 (0.86)	77.4	260		127 (0.30)		
	4[n]	55.9 (22)	83.8 (33)	3,364 (7,400)	3.3 (10.8)	0.25	871 (1,600)		364 (800)	2.6	7.2	31,820 (13,680)			118 (260)	2.2	400		267 (0.63)	83.5	280		127 (0.30)		

TABLE 84 (continued)

								Coal					Sorbent				Flue gas								
Test number	Test condition number	Static bed depth cm (in.)	Estimated[a] expanded bed depth cm (in.)	Air feed rate kg/hr (lb/hr)	Calculated[b] gas velocity m/s (ft/s)	Estimated[a] gas residence time (sec)	Bed temperature °C (°F)	Type	Feed rate kg/hr (lb/hr)	% S	% Ash	HHV kJ/kg (Btu/lb)	Type	Size μm	Feed rate kg/hr (lb/hr)	Ca/S ratio	SO_2 IRA[c] ppm	SO_2 wet chemical analysis[d] ppm	SO_2 ng/J[e] (lb/10^6 Btu)	SO_2 reduction[f] %	NO IRA[c] ppm	NO_x PDS[g] ppm	NO_x ng/J[h] (lb/10^6 Btu)	Fly ash[f] kg/hr (lb/hr)	Fly ash[f] ng/J[h] (lb/10^6 Btu)
25	1	61.0 (24)	91 (36)	3,364 (7,400)	3.2 (10.6)	0.28	840 (1,550)	Ohio #8 Seam Unwashed	373 (820)	4.5	10.7	30,084 (12,934)				0	3,750	3,840	3,009 (7.0)	0	240	305	138 (0.32)	3.5 (7.8)	318 (0.74)
	2	61.0 (24)	91 (36)	3,364 (7,400)	3.2 (10.6)	0.28	840 (1,550)		382 (840)	4.5	10.7	30,084 (12,934)	1337R	-44	172 (378)	1.7	1,100	1,210	858 (2.0)	72.5	240		138 (0.32)	6.5 (14.3)	567 (1.32)
	3	61.0 (24)	91 (36)	3,364 (7,400)	3.2 (10.4)	0.29	827 (1,520)		382 (840)	4.5	10.7	30,084 (12,934)			172 (378)	1.7	950		776 (1.8)	74.2	240		138 (0.32)		
26	1	50.8 (20)	76 (30)	3,364 (7,400)	3.4 (11.1)	0.23	904 (1,660)	Ohio #8 Seam Unwashed	364 (800)	4.5	10.7	30,084 (12,934)				0	3,750	3,740	3,009 (7.0)	0	220		127 (0.30)	5.6 (12.4)	516 (1.20)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.4 (11.1)	0.23	904 (1,660)		364 (800)	4.5	10.7	30,084 (12,934)	1337R	-44	164 (360)	1.7	1,350	1,350	1,077 (2.5)	64.2	220		127 (0.30)	7.6 (16.8)	696 (1.62)
	3	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.7)	0.23	860 (1,580)		364 (800)	4.5	10.7	30,084 (12,934)			182 (400)	1.9	1,100	1,160	876 (2.0)	70.9	220		127 (0.30)		
27	1	50.8 (20)	76 (30)	3,364 (7,400)	3.4 (11.0)	0.23	888 (1,630)	Ohio #8 Seam Unwashed	348 (765)	4.5	10.7	30,084 (12,934)				0	3,700	3,680	3,009 (7.0)	0	320	320	187 (0.44)	4.8 (10.5)	456 (1.06)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.7)	0.23	854 (1,570)		348 (765)	4.5	10.7	30,084 (12,934)	1359R	-44	100 (220)	2.0	950	980	782 (1.8)	74.0	320		187 (0.44)	7.5 (16.5)	718 (1.67)
28	1	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)	Ohio #8 Seam Washed	339 (745)	2.6	7.2	31,820 (13,680)				0	2,850	2,820	1,634 (3.8)	0	260		107 (0.25)	4.0 (8.9)	374 (0.87)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)		339 (745)	2.6	7.2	31,820 (13,680)	1359R	-44	68 (150)	2.4	800	810	464 (1.1)	71.6	260		107 (0.25)	5.6 (12.4)	525 (1.22)
	3	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)		361 (795)	2.6	7.2	31,820 (13,680)			68 (150)	2.2	950	1,020	574 (1.3)	64.9	260		107 (0.25)		
29	1	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)	Ohio #8 Seam Unwashed	327 (720)	4.5	10.7	30,084 (12,934)				0	3,770	3,730	3,009 (7.0)	0	240	300	138 (0.32)	5.5 (12.1)	559 (1.30)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)		327 (720)	4.5	10.7	30,084 (12,934)	1359R	-44	80 (175)	1.7	1,500	1,480	1,204 (2.8)	60.0	240		138 (0.32)		
	3	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)		327 (720)	4.5	10.7	30,084 (12,934)			100 (220)	2.0	1,000	1,080	797 (1.85)	73.5	240		138 (0.32)	6.7 (14.7)	679 (1.58)
	4[h]	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.8)	0.23	871 (1,600)		327 (720)	4.5	10.7	30,084 (12,934)			100 (220)	2.0	1,000		797 (1.85)	73.5	240		138 (0.32)		

TABLE 84 (continued)

Test number	Test condition number	Static bed depth cm (in.)	Estimated[a] expanded bed depth cm (in.)	Air feed rate kg/hr (lb/hr)	Calculated[b] gas velocity m/s (ft/s)	Estimated[a] gas residence time (sec)	Bed temperature °C (°F)	Coal					Sorbent				Flue gas							Fly ash[r] kg/hr (lb/hr)	Fly ash[r] ng/J[h] (lb/10⁶ Btu)
								Type	Feed rate kg/hr (lb/hr)	% S	% Ash	HHV kJ/kg (Btu/lb)	Type	Size μm	Feed rate kg/hr (lb/hr)	Ca/S ratio	SO_2 IRA[c] ppm	SO_2 wet chemical analysis[d] ppm	SO_2 ng/J[e] (lb/10⁶ Btu)	SO_2 reduction[f] %	NO IRA[c] ppm	NO_x PDS[g] ppm	NO_x ng/J[h] (lb/10⁶ Btu)		
30	1	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	882 (1,620)	Ohio #8 Seam Washed	336 (740)	2.6	7.2	31,820 (13,680)				0	2,570	2,650	1,636 (3.8)	0	250	285	116 (0.27)	3.5 (7.7)	327 (0.76)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	882 (1,620)		345 (760)			31,820 (13,680)	1359H	-44	30 (66)	1.6	1,290	1,360	917 (1.9)	50.0	250		116 (0.27)	5.2 (11.4)	473 (1.10)
	3	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	882 (1,620)		345 (760)			31,820 (13,680)			40 (88)	1.8	1,020	1,030	667 (1.5)	60.6	250		116 (0.27)		
31	1	50.8 (20)	76 (30)	3,455 (7,600)	3.4 (11.2)	0.22	882 (1,620)	Ohio #8 Seam Washed	373 (820)	2.6	7.2	31,820 (13,680)				0	2,600	2,630	1,636 (3.8)	0	270	305	122 (0.28)	5.1 (11.2)	430 (1.0)
	2	50.8 (20)	76 (30)	3,455 (7,600)	3.4 (11.2)	0.22	882 (1,620)		350 (770)			31,820 (13,680)	1359H	-44	29 (63)	1.3	1,200	1,250	755 (1.76)	53.8	270		122 (0.28)	5.0 (10.9)	467 (1.04)
	3	50.8 (20)	76 (30)	3,455 (7,600)	3.4 (11.2)	0.22	882 (1,620)		364 (800)			31,820 (13,680)			36 (80)	1.6	1,000	1,080	629 (1.5)	61.5	270		122 (0.28)		
32	1	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	877 (1,610)	Ohio #8 Seam Washed	318 (700)	2.6	7.2	31,820 (13,680)				0	2,650	2,690	1,636 (3.8)	0	290	310	129 (0.30)	5.0 (10.9)	490 (1.14)
	2	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	877 (1,610)		327 (720)			31,820 (13,680)	1359R	-44	44 (97)	1.6	1,020	1,050	623 (1.44)	61.9	290		129 (0.30)	6.2 (13.7)	598 (1.39)
	3	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	877 (1,610)		327 (720)			31,820 (13,680)			49 (108)	1.8	970	970	576 (1.33)	64.9	290		129 (0.30)		
	4[n]	50.8 (20)	76 (30)	3,364 (7,400)	3.3 (10.9)	0.23	877 (1,610)		327 (720)			31,820 (13,680)			49 (108)	1.8	780		482 (1.1)	70.5	290		129 (0.30)		

[a] Estimated by GCA.

[b] Calculated by GCA.

[c] Infrared analysis.

[d] Similar to EPA Reference Method 6.

[e] Estimated by GCA, based on IR results.

[f] Based on IR results.

[g] Similar to EPA Reference Method 7.

[h] Estimated by GCA.

[i] Test of effect of varying excess air.

[j] Coal transition from East Kentucky to Ohio No. 8.

[k] 22.7 kg (50 lb) limestone added as slug.

[l] R = raw limestone.

[m] Gas sample system leak.

[n] Fly ash recirculation during test.

[o] Natural mine limestone - 72 percent $CaCO_3$.

[p] Steam added to air inlet.

[q] Hydrated limestone.

[r] Particulate measured using isokinetic probe system (see Subsection 7.3.4).

Note: NR = Not reported.

TABLE 85. PER-FBM EMISSION SOURCE TEST DATA RECORDED IN TESTS CONDUCTED THROUGH 1975 WITH SEWICKLEY COAL[8]

Test number	Operating bed depth cm (in.)	Superficial gas velocity m/sec (ft/sec)	Bed temperature ^{o}C (^{o}F)	Gas residence time (sec)	Coal			Limestone				SO_2	
					Type	Feed rate kg/hr (lb/hr)	S %	Type	Size*	Feed rate kg/hr (lb/hr)	Ca/S ratio	ppm†	ng/J ($lb/10^6$ Btu)
636	86.1 (33.9)	4.3 (14)	815 (1500)	0.202	Sewickley	336 (740)	4.1 - 4.5	Germany Valley		216 (475)	4.4	650	679 (1.58)
637-1	102.9 (40.5)	3.8 (12.4)	815 (1500)	0.272	Sewickley	320 (705)	4.1 - 4.5	Greer		170 (374)	2.9	500	598 (1.39)
637-2	101.3 (39.9)	3.8 (12.5)	827 (1520)	0.266	Sewickley	320 (705)	4.1 - 4.5	Greer		189 (416)	3.2	370	512 (1.19)
639	96.8 (38.1)	4.7 (15.5)	857 (1575)	0.205	Sewickley	350 (770)	4.1 - 4.5	Greer		202 (445)	3.5	490	473 (1.10)
621	94.0 (37)	4.5 (14.6)	815 (1500)	0.211	Sewickley	334 (735)	4.1 - 4.5	Greer		133 (292)	2.9	1,200	1,071 (2.49)
630	96.5 (38)	3.8 (12.7)	815 (1500)	0.249	Sewickley	306 (674)	4.1 - 4.5	Germany Valley		114 (251)	2.76	1,120	967 (2.25)

*Size ranged from 370 to 4,760 μm.

†By IR analyzer.

TABLE 86. OPERATING CONDITIONS AND RESULTS OF FLUIDYNE 500-HR TEST IN 3.3 FT × 5.3 FT VERTICLE SLICE COMBUSTOR[10]

OPERATING CONDITIONS

• Fuel Characteristics	
1. Type	Illinois No. 6
2. Surface Moisture (%)	2 - 11
3. Feed rate	68 - 227 kg/hr (150 - 500 lb/hr)
4. % Sulfur	3.6
5. Feed location	In-bed
• Sorbent Characteristics	
1. Type	Owatonna Dolomite - 1/4 in.
2. Surface Moisture (%)	3 - 7
3. Ca/S	1.1 - 2.2
4. Feed rate	23 - 82 kg/hr (50 - 180 lb/hr)
5. Feed location	In-bed
• Bed Temperature	718° to 796°C (1325° to 1465°F)
• Bed Depth	1.1 - 1.2 m (42 - 47 in.)
• Superficial Velocity	0.6 - 1.3 m/sec (2.0 - 4.2 ft/sec)
• Flue Gas Excess Air Level	30 - 130 percent
• Process Air Flow Rate	454 - 5675 kg/hr (1,000 - 12,500 lb/hr)
• Total Heat Output	0.5 - 1.6 MW_t (1.65 - 5.5 × 10^6 Btu/hr)
• Recycle of Elutriated Particulates	Yes
• Combustion Efficiency (%)	93.5 to 96.3

RESULTS OF TESTING

Load	Low	High
Bed Temperature, °C (°F)	718 (1325)	796 (1465)
Superficial Velocity, m/sec (ft/sec)	0.76 (2.5)	1.1 (3.6)
Kg Dolomite/Kg Coal	0.46	0.31
Ca/S Ratio*	2.4	1.7
SO_2 Control Efficiency (%)†	80	80
NO_x emission ng/J (lb/10^6 Btu)	236 (0.55)	159 (0.37)
Excess Air (%)	130	30

*Estimated by GCA from kg dolomite/kg coal.

†Estimated by GCA from coal heating value and sulfur content and SO_2 outlet level of 516 ng/J (1.2 lb/10^6 Btu).

TABLE 87. OPERATING CONDITIONS AND RESULTS OF FLUIDYNE RUN 35 IN 3.3 FT × 5.3 FT VERTICAL SLICE COMBUSTOR[11]

OPERATING CONDITIONS	
● Fuel Characteristics	
1. Type	Illinois No. 6
2. Feed rate	173 kg/hr (380 lb/hr)
3. % Sulfur	3.6
4. Feed location	Above-bed
● Sorbent Characteristics	
1. Type	Owatonna Dolomite
2. Ca/S ratio	2.38
3. Feed rate	77 kg/hr (170 lb/hr)
4. Feed location	Above-bed
● Bed Temperature	772°C (1421°F)
● Bed Depth	1.1 m (45 in.)
● Superficial velocity	1 m/sec (3.21 ft/sec)
● Excess air	50 percent
● Recycle of Elutriated Particulate	Yes
● Gas Phase Residence Time*	0.86 sec
RESULTS OF TESTING	
Ca/S ratio	2.38
SO_2 Control Efficiency (%)	87.2

*Estimated by GCA.

TABLE 88. EMISSION SOURCE TEST DATA: NCB 6-IN. DIAMETER FBC UNIT FIRING WELBECK, PARK HILL, ILLINOIS, AND PITTSBURGH COALS WITH U.K. LIMESTONE AT A TEMPERATURE OF 799°C (1470°F)[12]

Test No.	Coal type	Fluidizing velocity, m/sec (ft/sec)	Oxygen in off gas, %	Ca/S molar feed ratio	SO_2, ppm	S retention, %	SO_2 reduction, %
1.5	Illinois	0.9 (3)	2.6	1.1	1785	55	55
1.6	Illinois	0.9 (3)	2.2	2.2	1000	66	67
1.7	Illinois	0.9 (3)	2.3	3.2	229	94	94
1.8	Illinois	0.9 (3)	2.6	0	3896	1	0
2.1*	Welbeck	0.6 (2)	3.1	0	1087	7	0
2.2*	Welbeck	0.6 (2)	3.1	0.8	488	58	55
2.3	Welbeck	0.6 (2)	3.1	0.8	513	56	52
2.4	Welbeck	0.9 (3)	3.1	0.8	599	54	45
2.5	Welbeck	0.9 (3)	2.3	1.8	207	82	81
2.6	Welbeck	0.9 (3)	2.2	2.9	33	98	97
2.7†	Welbeck	0.9 (3)	0.3	0.8	612	61	48
2.8	Welbeck	0.6 (2)	2.6	1.8	158	86	85
2.9	Park Hill	0.9 (3)	2.0	0.5	1690	15	—
2.10	Park Hill	0.9 (3)	2.5	1.3	1045	30	25
2.11	Park Hill	0.9 (3)	2.0	2.7	237	79	78
2.12	Park Hill	0.9 (3)	2.3	0	2091	12	0
2.13	Park Hill	0.9 (3)	2.1	1.1	1210	49	42
2.14	Park Hill	0.9 (3)	2.5	2.6	322	86	84
2.15*	Park Hill	0.9 (3)	2.0	0.8	547	54	50
4.1	Pittsburgh	0.9 (3)	2.3	0	1980	5	0
4.2	Pittsburgh	0.9 (3)	2.4	1.0	1372	41	31
4.3	Pittsburgh	0.9 (3)	2.2	2.1	938	57	52
4.4	Pittsburgh	0.9 (3)	2.4	3.1	241	90	88
5.1	Welbeck	0.9 (3)	2.8	1.6	342	79	71

*Tests with primary fines recycle.

†Test at substiochiometric conditions.

TABLE 89. EMISSION SOURCE TEST DATA: NCB 6 INCH DIAMETER FBC UNIT FIRING ILLINOIS COAL WITH LIMESTONE 1359 AT A FLUIDIZING VELOCITY OF 0.9 m/sec (3 ft/sec)[12]

Test No.	Bed temperature °C (°F)	Bed depth, m (ft)	Oxygen in off gas, %	Ca/S molar feed ratio	SO_2, ppm	S retention, %	SO_2 reduction, &
1.1	799 (1470)	0.6 (2)	2.9	0	4023	0	0
1.2	799 (1470)	0.6 (2)	2.8	1.5	2118	47	47
1.3	799 (1470)	0.6 (2)	2.6	2.2	1450	64	63
1.4	799 (1470)	0.6 (2)	2.9	3.3	680	78	78
3.1	699 (1290)	0.6 (2)	2.7	1.1	3376	24	15
3.2	699 (1290)	0.6 (2)	2.4	2.2	3245	26	18
3.3	799 (1470)	0.9 (3)	2.7	1.1	1930	49	51
3.4	799 (1470)	0.9 (3)	2.6	2.1	1136	70	72
3.5*	799 (1470)	0.6 (2)	2.4	1.1	1523	61	61
3.6*	799 (1470)	0.6 (2)	2.5	3.6	278	92	93

*Tests with -125 μm limestone particles.

TABLE 90. EMISSION SOURCE TEST DATA: NCB 6 INCH DIAMETER FBC UNIT FIRING PITTSBURGH AND WELBECK COALS WITH LIMESTONE 18 AT A TEMPERATURE OF 799°C (1470°F), BED DEPTH OF 0.6 m (2 feet) AND FLUIDIZING VELOCITY OF 0.9 m/sec (3 ft/sec)[12]

Test No.	Coal type	Oxygen in off gas, %	Ca/S molar feed ratio	SO_2, ppm	S retention, %	SO_2 reduction, %
4.1	Pittsburgh	2.3	0	1980	5	0
4.5	Pittsburgh	2.5	0.9	1137	50	43
4.6	Pittsburgh	2.2	1.7	581	75	71
4.7	Pittsburgh	2.3	2.6	185	92	91
4.8*	Pittsburgh	2.1	0.9	1238	55	38
4.9†	Pittsburgh	2.5	0.9	1115	60	43
5.2	Welbeck	2.6	1.9	236	85	80

*Test with lime rich bed.

†Test with shale bed.

TABLE 91. EMISSION TEST DATA MEASURED FROM ANL'S 6-IN. AFBC UNIT[13-16]

	Test conditions				Fuel characteristics					Sorbent characteristics				Emission characteristics				
														SO_2			NO_x	
Test No.	Bed temp. °C (°F)	Superficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas residence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/s (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention %	ppm	ng/J (lb/10^6 Btu)
CC-1-1	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	dolomite 1337	—	0.0 (0.0)	0.0	1350	3295 (7.66)	0.0	—	—
CC-1-2	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	300	—	3.5	450	1090 (2.53)	67.0	—	—
CC-1-3	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	300	—	5.1	350	855 (1.99)	74.0	—	—
CC-2-1	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	—	0.0 (0.0)	0.0	2000	3295 (7.66)	0.0	—	—
CC-2-2	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	100	—	1.4	1150	1910 (4.44)	42.0	—	—
CC-2-3	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	100	—	2.8	500	825 (1.92)	75.0	—	—
CC-3-1	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	—	0.0 (0.0)	0.0	1350	3295 (7.66)	0.0	—	—
CC-3-2	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	300	—	2.0	600	1290 (3.00)	61.0	—	—
CC-3-3	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	300	—	2.9	400	855 (1.99)	74.0	—	—
CC-3-4	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	300	—	4.0	200	430 (1.00)	87.0	—	—
CC-4-1	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	—	0.0 (0.0)	0.0	2250	3295 (7.66)	0.0	—	—
CC-4-2	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	100	—	1.5	1600	2340 (5.44)	29.0	—	—
CC-4-3	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	100	—	2.1	1100	1610 (3.75)	51.0	—	—
CC-4-4	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	100	—	2.6	900	1315 (3.06)	60.0	—	—
CC-7-1	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1337	—	0.0 (0.0)	0.0	2850	3295 (7.66)	0.0	—	—
CC-7-2	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	3.7 (29.7)	limestone 1360	1200	0.2 (1.5)	2.5	800	920 (2.14)	72.0	—	—
CC-9	870 (1600)	1.14 (3.75)	0.38 (15)	0.33	28,126 (12,092)	4.63	1.31	12.39	—	1360	1200	—	4.2	400	460 (1.07)	86.0	—	—
SACC-1	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,126 (12,092)	4.63	1.31	12.39	3.7 (29.1)	limestone 1359	25	1.2 (9.69)	1.9	1300	1055 (2.45)	68.0	440	255 (0.59)

TABLE 91 (continued)

Test No.	Test conditions				Fuel characteristics					Sorbent characteristics				Emission characteristics				
														SO_2			NO_x	
	Bed temp. °C (°F)	Super-ficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas resi-dence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 reten-tion %	ppm	ng/J (lb/10^6 Btu
SACC-2	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	—	1359	600	—	2.4	1600	1600 (3.72)	53.0	400	230 (0.53)
SACC-3	870 (1600	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	3.8 (30.2)	1359	25	1.2 (9.82)	2.6	1700	1630 (3.79)	52.0	400	230 (0.53)
SACC-4	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	3.7 (29.7)	1359	25	1.2 (9.79)	2.2	360	445 (1.03)	87.0	360	205 (0.48)
SA1A	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	—	1359	25	—	0.0	3780	3400 (7.91)	0.0	720	415 (0.96)
SA1B	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	—	1359	25	—	1.5	1600	1425 (3.32)	58.0	600	345 (0.80)
SA1C	843 (1500)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	—	1359	25	—	2.0	1250	1120 (2.61)	67.0	600	345 (0.40)
SA1D	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	—	1359	25	—	2.6	650	574 (1.34)	83.0	600	345 (0.80)
SA1E	843 (1500)	0.91 (3.0)	0.61 (24)	0.67	28.482 (12,245)	4.84	1.11	13.13	—	1359	25	—	4.2	400	375 (0.87)	89.0	650	375 (0.87)
SACC-5	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,482 (12,245)	4.84	1.11	13.13	—	1359	25	—	2.2	800	715 (1.66)	79.0	420	240 (0.56)
SACC-5R	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	2.2	800	715 (1.66)	79.0	420	240 (0.56)
SACC9-1	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.7	1400	1225 (2.85)	64.0	500	290 (0.67)
SACC9-2	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.2	2290	2005 (4.66)	41.0	400	230 (0.53)
SACC9-3	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.7	1380	1190 (2.77)	65.0	460	260 (0.61)
SACC9-5	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	3.0	800	715 (1.66)	79.0	500	290 (0.67)
SACC6-1	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.3	2000	1665 (3.87)	51.0	420	240 (0.56)
SACC6-2	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.7	1650	1395 (3.24)	59.0	400	230 (0.53)
SACC-7	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	1400	—	1.6	3350	2785 (6.48)	18.0	550	315 (0.73)
SACC8-1	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.0	2400	2070 (4.82)	39.0	520	295 (0.69)

TABLE 91 (continued)

	Test conditions				Fuel characteristics				Sorbent characteristics					Emission characteristics				
														SO_2			NO_x	
Test No.	Bed temp. °C (°F)	Superficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas residence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention %	ppm	ng/J (lb/10^6 Btu)
SACC8R-2	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.0	2650	2280 (5.30)	33.0	510	290 (0.68)
SACC8R-3	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	2.4	1150	985 (2.29)	71.0	510	290 (0.68)
SACC8-4	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	2.4	1350	1055 (2.45)	69.0	470	270 (0.63)
SACC8R-5	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	2.4	1550	1225 (2.85)	64.0	420	240 (0.56)
SA2A	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	0.0	4400	3400 (7.91)	0.0	800	460 (1.07)
SA2B	870 (1600)	2.7 (9.0)	0.61 (24)	0.22	28,475 (12,242)	4.84	1.11	13.13	—	1359	25	—	1.5	2050	1565 (3.64)	54.0	560	320 (0.75)
SA2C	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475) (12,242)	4.84	1.31	13.13	—	1359	25	—	2.6	1150	885 (2.06)	74.0	600	345 (0.80)
SA2D	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475) (12,242)	4.84	1.31	13.13	—	1359	25	—	3.0	850	645 (1.50)	81.0	600	345 (0.80)
SA2E	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	25	—	3.7	620	475 (1.11)	86.0	600	345 (0.80)
SA3A	900 (1650)	0.91 (3.0)	0.61 (3.0)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	0.0	4000	3400 (7.91)	0.0	—	—
SA3B	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	2.4	1600	1360 (3.16)	60.0	—	—
SA3C	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	0.8	1800	1530 (3.56)	55.0	—	—
SA3D	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475) (12,242)	4.84	1.31	13.13	—	1359	103	—	0.6	2300	1975 (4.59)	42.0	—	—
SA3E	900 (1650)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	0.8	2300	1975 (4.59)	42.0	—	—
SA4A	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	0.0	3400	3400 (7.91)	0.0	760	435 (1.01)
SA-3	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	4.0	160	170 (0.40)	95.0	550	315 (0.73)
SA4C	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	1.7	1300	1395 (3.24)	59.0	570	325 (0.76)
SA4D	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	0.6	2400	2315 (5.38)	32.0	600	345 (0.80)

TABLE 91 (continued)

	Test conditions				Fuel characteristics				Sorbent characteristics					Emission characteristics				
														SO_2			NO_x	
Test No.	Bed temp. °C (°F)	Super-ficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas resi-dence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 reten-tion %	ppm	ng/J (lb/10^6 Btu)
SA4E	843 (1550)	0.91 (3.0)	0.61 (24)	0.67	28,475 (12,242)	4.84	1.31	13.13	—	1359	103	—	2.4	920	915 (2.13)	73.0	530	305 (0.71)
BC-1-1	870 (1600)	0.91 (3.0)	0.61 (24)	0.67	28,475) (12,242)	4.84	1.31	13.13	—	1359	25	—	1.2	—	1600 (3.72)	53.0	—	—
BC-1-2	870 (1600)	0.87 (2.85)	0.61 (24)	0.70	28,475 (12,242)	4.84	1.31	13.13	—	1359	25	—	2.0	—	1190 (2.77)	65.0	—	—
BC-2	870 (1600)	0.87 (2.85)	0.61 (24)	0.70	28.475 (12,242)	4.84	1.31	13.13	—	Tymoch-tee	575	—	1.6	530	445 (1.03)	87.0	500	290 (0.67)
BC-3	870 (1600)	0.87 (2.85)	0.61 (24)	0.70	28,475 (12,242)	4.84	1.31	13.13	—	Tymoch-tee	575	—	1.5	850	750 (1.74)	78.0	550	315 (0.73)
BC-4	870 (1600)	0.87 (2.85)	0.61 (24)	0.70	28,475 (12,242)	4.84	1.31	13.13	—	Tymoch-tee	44	—	0.6	1850	1870 (4.35)	45.0	350	200 (0.47)
BC-4-1	870 (1600)	0.87 (2.85)	0.61 (24)	0.70	28,475 (12,242)	4.84	1.31	13.13	—	Tymoch-tee	44	—	1.2	1050	1055 (2.45)	69.0	395	230 (0.53)
BC-5-1	870 (1600)	0.87 (2.85)	0.61 (24)	0.70	28,475 (12,242)	4.84	1.31	13.13	—	Tymoch-tee	44	—	1.5	1250	1155 (2.69)	66.0	365	210 (0.49)
BC-5-2	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	—	Tymoch-tee	44	—	2.1	620	575 (1.34)	83.0	400	230 (0.53)
BC-6-1	816 (1500)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.4)	1359	615	—	—	130	140 (0.32)	96.0	600	345 (0.80)
BC-6-2	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.4)	1359	615	—	—	250	1800 (4.19)	47.0	400	230 (0.53)
BC-6-3	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.4)	1359	615	0.2 (1.6)	2.6	960	715 (1.66)	79.0	380	220 (0.51)
BC-7-1	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.9)	1359	615	0.0 (0.0)	0.0	2250	1530 (3.56)	55.0	340	195 (0.45)
BC-7-2	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.7 (5.2)	1359	615	0.2 (1.8)	2.3	930	575 (1.34)	83.0	220	125 (0.29)
BC-8-1	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.5)	1359	615	0.0 (0.0)	0.0	3350	2480 (5.77)	27.0	310	175 (0.41)
BC-8-2	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.5 (4.3)	1359	615	0.2 (1.6)	2.5	940	715 (1.66)	79.0	320	185 (0.43)
BC-9	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.5 (4.1)	1359	630	0.2 (1.7)	2.3	1100	885 (2.06)	74.0	400	230 (0.53)
BC-10-A	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.5 (4.3)	1337	540	0.3 (2.6)	2.2	910	645 (1.50)	81.0	440	255 (0.59)

TABLE 91 (continued)

	Test conditions				Fuel characteristics					Sorbent characteristics				Emission characteristics				
														SO_2			NO_x	
Test No.	Bed temp. °C (°F)	Superficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas residence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention %	ppm	ng/J (lb/10^6 Btu)
BC-10-B	788 (1450)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.5 (4.3)	1337	540	0.3 (2.6)	2.2	470	305 (0.71)	91.0	340	195 (0.45)
BC-10-C	982 (1800)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.5 (4.3)	1337	540	0.3 (2.6)	2.2	3650	2820 (6.56)	17.0	460	260 (0.61)
AR-1-A	760 (1400)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.6)	1359	490	0.2 (1.7)	2.5	2400	1905 (4.43)	44.0	250	140 (0.33)
AR-1-B	788 (1450)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.6)	1359	490	0.2 (1.7)	2.5	1460	1190 (2.77)	65.0	280	160 (0.37)
AR-1-C	816 (1500)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.6)	1359	490	0.2 (1.7)	2.5	420	305 (0.71)	91.0	360	205 (0.48)
AR-1-D	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.6)	1359	490	0.2 (1.7)	2.5	420	305 (0.71)	91.0	430	245 (0.57)
AR-1-E	870 (1600)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.6)	1359	490	0.2 (1.7)	2.5	900	475 (1.11)	86.0	430	245 (0.57)
AR-1-F	760 (1400)	0.85 (2.8)	0.61 (24)	0.71	28,475 (12,242)	4.84	1.31	13.13	0.6 (4.6)	1359	490	0.2 (1.7)	2.5	2450	1680 (3.72)	53.0	270	155 (0.36)
BRIT-1	800 (1470)	0.76 (2.5)	0.61 (24)	0.71	27,463 (11,807)	1.28	1.21	18.07	0.6 (5.0)	B-SONK	440	0.1 (0.42)	2.2	320	205 (0.48)	78.0	350	200 (0.47)
BRIT-2	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	27,463 (11,807)	1.28	1.21	18.07	0.6 (4.9)	B-SONK	440	0.1 (0.69)	3.65	250	170 (0.39)	82.0	310	175 (0.41)
BRIT-3	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	27,463 (11,807)	1.28	1.21	18.07	0.7 (5.2)	B-SONK	440	0.1 (0.23)	1.2	660	430 (1.00)	55.0	265	150 (0.35)
AMER-1	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	28,290 (12,163)	4.14	1.18	12.08	0.6 (4.5)	1359	555	0.1 (0.53)	1.05	2480	1815 (4.22)	38.0	240	140 (0.32)
AMER-2	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	28,290 (12,163)	4.14	1.18	12.08	0.6 (4.6)	1359	555	0.2 (1.6)	2.9	870	645 (1.50)	78.0	260	150 (0.35)
AMER-3	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	28,290 (12,163)	4.14	1.18	12.08	0.6 (4.6)	1359	555	0.1 (1.05)	1.95	1460	1085 (2.52)	63.0	215	125 (0.29)
AMER-33	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	28,290 (12,163)	4.14	1.18	12.08	0.6 (4.6)	1359	550	0.2 (1.5)	2.75	840	615 (1.43)	79.0	240	140 (0.32)
AM-BRIT	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	28,290 (12,163)	4.14	1.18	12.08	0.6 (4.6)	B-SONK	440	0.1 (1.0)	1.9	1300	940 (2.18)	68.0	250	140 (0.33)
BRIT-AM	800 (1470)	0.79 (2.6)	0.61 (24)	0.77	27,463 (11,807)	1.28	1.21	18.07	0.7 (5.2)	1359	555	0.1 (0.38)	1.9	500	320 (0.74)	66.0	265	150 (0.35)
AR2A	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.3)	1359	490	0.2 (1.6)	2.6	470	–	–	390	225 (0.52)

TABLE 91 (continued)

Test No.	Test conditions				Fuel characteristics					Sorbent characteristics				Emission characteristics				
														SO_2			NO_x	
	Bed temp. °C (°F)	Superficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas residence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size µm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention %	ppm	ng/J (lb/10^6 Btu)
AR2B	843 (1550)	0.76 (2.5)	0.61 (24)	0.80	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.4)	1359	490	0.2 (1.7)	2.6	730	–	–	350	200 (0.47)
AR2C	843 (1550)	0.73 (2.4)	0.61 (24)	0.83	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.5)	1359	490	0.2 (1.7)	2.6	1250	–	–	300	170 (0.40)
AR2D	843 (1550)	0.94 (3.1)	0.61 (24)	0.65	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.5)	1359	490	0.2 (1.8)	2.6	850	–	–	430	245 (0.57)
AR4	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.2)	1359	490	0.2 (1.6)	2.8	750	–	–	310	175 (0.41)
AR5A	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.5 (3.8)	1359	490	0.2 (1.6)	3.0	1100	–	–	370	210 (0.49)
AR5B	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.4)	1359	490	0.4 (3.1)	5.5	200	–	–	350	200 (0.47)
AR5C	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.1)	1359	490	0.3 (2.4)	4.6	160	–	–	440	255 (0.59)
AR5D	843 (1550)	0.85 (2.8)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.1)	1359	490	0.2 (1.3)	2.5	720	–	–	330	190 (0.44)
AR6C	843 (1550)	2.26 (7.4)	0.61 (24)	0.27	28,290 (12,163)	3.7	1.18	10.85	0.5 (5.6)	1359	1640	0.4 (3.5)	4.2	1500	–	–	470	270 (0.63)
AMER6D	844 (1552)	0.80 (2.64)	0.61 (24)	0.76	28,290 (12,163)	3.7	1.18	10.85	0.7 (5.2)	1359	609	0.2 (1.8)	2.99	1516	865 (2.01)	67	214	125 (0.29)
AMER6E	840 (1544)	0.80 (2.64)	0.61 (24)	0.76	28,290 (12,163)	3.7	1.18	10.85	0.7 (5.4)	1359	609	0.2 (1.8)	2.94	1195	76 (1.76)	71	264	150 (0.35)
AMER8A	841 (1545)	0.87 (2.85)	0.36 (14)	0.41	28,290 (12,163)	3.7	1.18	10.85	0.7 (5.0)	1359	–	0.3 (2.3)	3.99	891	–	–	348	200 (0.46)
AMER8B	849 (1560)	0.91 (2.98)	0.61 (24)	0.67	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.8)	1359	–	0.3 (2.4)	4.28	751	–	–	299	160 (0.37)
AMER8C	845 (1553)	0.91 (3.15)	1.17 (46)	1.22	28,290 (12,163)	3.7	1.18	10.85	0.6 (5.0)	1359	–	0.2 (2.3)	3.98	570	–	–	352	200 (0.47)
HUMP1A	783 (1441)	0.79 (2.60)	0.61 (24)	0.77	–	2.4	–	–	0.6 (4.1)	1359	–	0.2 (1.3)	4.18	25	–	–	464	265 (0.62)
HUMP1B	842 (1548)	0.84 (2.77)	0.61 (24)	0.72	–	2.4	–	–	0.5 (4.2)	1359	–	0.2 (1.4)	4.30	380	–	–	529	305 (0.71)
HUMP1C	900 (1650)	0.86 (2.83)	0.61 (24)	0.71	–	2.4	–	–	0.5 (4.1)	1359	–	0.2 (1.4)	4.45	980	–	–	610	350 (0.81)
HUMP1D	791 (1455)	0.80 (2.62)	0.61 (24)	0.76	–	2.4	–	–	0.5 (3.8)	1359	–	0.2 (1.3)	4.58	64	–	–	529	305 (0.71)

TABLE 91 (continued)

	Test conditions				Fuel characteristics					Sorbent characteristics				Emission characteristics				
														SO_2			NO_x	
Test No.	Bed temp. °C (°F)	Superficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas residence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention %	ppm	ng/J (lb/10^6 Btu)
HUMP-1E	757 (1395)	0.74 (2.42)	0.61 (24)	0.83	–	2.4	–	–	0.5 (4.0)	1359	–	0.2 (1.3)	4.44	276	–	–	447	250 (0.50)
HUMP2A2	791 (1456)	0.80 (2.61)	0.61 (24)	0.77	–	2.4	–	–	0.5 (4.0)	1359	–	0.1 (0.8)	2.67	564	–	–	336	195 (0.45)
HUMP2B3	784 (1443)	0.78 (2.57)	0.61 (24)	0.78	–	2.4	–	–	0.5 (4.1)	1359	–	0.1 (0.3)	1.00	1310	–	–	461	260 (0.61)
HUMP3	789 (1452)	0.77 (2.53)	0.61 (24)	0.79	–	2.4	–	–	0.5 (4.0)	1359	–	0.1 (0.3)	1.10	1500	–	–	486	280 (0.65)
HUMP3-2	786 (1446)	0.77 (2.52)	0.61 (24)	0.79	–	2.4	–	–	0.5 (4.0)	1359	–	0.1 (0.4)	1.28	1526	–	–	534	305 (0.71)
HUMP4-1	796 (1464)	0.78 (2.55)	0.61 (24)	0.78	–	2.4	–	–	0.5 (4.0)	1359	–	0.1 (0.3)	0.94	1571	–	47	531	305 (0.71)
HUMP4-2	783 (1441)	0.77 (2.51)	0.61 (24)	0.80	–	2.4	–	–	0.5 (4.0)	1359	–	0.1 (0.3)	0.94	1480	–	44	506	290 (0.67)
HUMP4-3	784 (1443)	0.76 (2.50)	0.61 (24)	0.80	–	2.4	–	–	0.5 (4.1)	1359	–	0.1 (0.3)	1.00	1413	–	42	433	250 (0.58)
HUMP4-4	787 (1448)	0.77 (2.52)	0.61 (24)	0.79	–	2.4	–	–	0.5 (4.0)	1359	–	0.1 (0.4)	1.46	1306	–	39	396	230 (0.53)
HP-5-A	718 (1325)	0.73 (2.40)	0.61 (24)	0.83	–	2.4	–	–	0.5 (4.0)	–	–	0.0 (0.0)	0.0	1910	–	0	462	265 (0.62)
HP-5-B	788 (1450)	0.85 (2.79)	0.61 (24)	0.72	–	2.4	–	–	0.5 (3.9)	–	–	0.0 (0.0)	0.0	1911	–	0	609	350 (0.81)
HP-5-C	837 (1538)	0.88 (2.90)	0.61 (24)	0.69	–	2.4	–	–	0.5 (3.6)	–	–	0.0 (0.0)	0.0	2051	–	0	626	355 (0.83)
HP-5-D	784 (1605)	0.91 (2.99)	0.61 (24)	0.67	–	2.4	–	–	0.5 (4.2)	–	–	0.0 (0.0)	0.0	2231	–	0	601	345 (0.80)
HP-5-E	718 (1325)	0.82 (2.70)	0.61 (24)	0.74	–	2.4	–	–	0.5 (4.3)	–	–	0.0 (0.0)	0.0	1987	–	0	585	335 (0.78)
HP6A	720 (1328)	0.79 (2.59)	0.61 (24)	0.77	–	2.4	–	–	0.6 (4.4)	–	–	0.0 (0.0)	0.0	2282	–	0	600	345 (0.80)
HP6B	782 (1439)	2.36 (7.73)	0.61 (24)	0.26	–	2.4	–	–	0.5 (4.0)	–	–	0.0 (0.0)	0.0	2119	–	0	684	390 (0.91)
HP6C	840 (1544)	0.87 (2.87)	0.61 (24)	0.70	–	2.4	–	–	0.6 (4.4)	–	–	0.0 (0.0)	0.0	2289	–	0	684	390 (0.91)
HP6D	894 (1642)	0.93 (3.04)	0.61 (24)	0.66	–	2.4	–	–	0.6 (5.0)	–	–	0.0 (0.0)	0.0	2452	–	0	642	370 (0.86)

TABLE 91 (continued)

	Test conditions				Fuel characteristics					Sorbent characteristics				Emission characteristics				
														SO_2			NO_x	
Test No.	Bed temp. °C (°F)	Superficial gas velocity m/s (ft/s)	Bed depth m (in.)	Gas residence time sec	Heating value kJ/kg (Btu/lb)	Sulfur %	Nitrogen %	Ash %	Feed rate g/S (lb/h)	Type	Mean size μm (in.)	Feed rate g/S (lb/h)	Ca/S ratio	ppm	ng/J (lb/10^6 Btu)	SO_2 retention %	ppm	ng/J (lb/10^6 Btu)
PBY2A	719 (1326)	0.73 (2.40)	0.61 (24)	0.83	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.4)	—	—	0.0 (0.0)	0.0	3903	2615 (6.08)	0	534	305 (0.71)
PBY2B	788 (1450)	0.79 (2.60)	0.61 (24)	0.77	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.1)	—	—	0.0 (0.0)	0.0	3677	2615 (6.08)	0	649	375 (0.87)
PBY2C	844 (1551)	0.82 (2.70)	0.61 (24)	0.74	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.2)	—	—	0.0 (0.0)	0.0	3759	2615 (6.08)	0	654	375 (0.87)
PBY2D	896 (1644)	0.85 (2.80)	0.61 (24)	0.71	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.3)	—	—	0.0 (0.0)	0.0	4095	2615 (6.08)	0	649	375 (0.87)
PBY2E	788 (1450)	0.79 (2.60)	0.61 (24)	0.77	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.2)	—	—	0.0 (0.0)	0.0	3733	2615 (6.08)	0	672	385 (0.90)
PEABY4	804 (1479)	0.80 (2.61)	0.61 (24)	0.77	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.0)	—	—	0.3 (2.1)			—	—	5	5 (0.007)
PEABY5	842 (1547)	0.82 (2.70)	0.61 (24)	0.74	28,290 (12,163)	3.7	1.18	10.85	0.5 (3.9)	1359	—	0.2 (1.2)	2.58	452	—	—	318	180 (0.42)
PBY5R	843 (1550)	0.82 (2.69)	0.61 (24)	0.74	28,290 (12,163)	3.7	1.18	10.85	0.5 (3.9)	1359	—	0.1 (1.1)	2.43	649	—	—	294	170 (0.39)
PEABY-6	844 (1551)	0.82 (2.68)	0.61 (24)	0.75	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.0)	1359	—	0.2 (1.3)	2.89	1169	—	—	388	225 (0.52)
PBY6R	843 (1550)	0.92 (3.01)	0.61 (24)	0.66	28,290 (12,163)	3.7	1.18	10.85	0.5 (3.9)	1359	—	0.1 (1.1)	2.43	845	—	—	236	135 (0.31)
AMER-333	799 (1471)	0.85 (2.79)	0.61 (24)	0.72	28,290 (12,163)	3.7	1.18	10.85	0.5 (4.3)	1359	—	0.2 (1.7)	3.50	660	—	—	121	70 (0.16)
AMER-333-3	798 (1468)	0.79 (2.59)	0.31 (12)	0.39	28,290 (12,163)	3.7	1.18	10.85	0.6 (4.7)	1359	—	0.2 (1.5)	2.75	1459	—	—	494	285 (0.66)
AMER-333-4	803 (1477)	0.78 (2.56)	0.31 (12)	0.39	28,290 (12,163)	3.7	1.18	10.85	0.6 (5.0)	1359	—	0.2 (1.6)	3.25	1172	—	—	203	115 (0.27)

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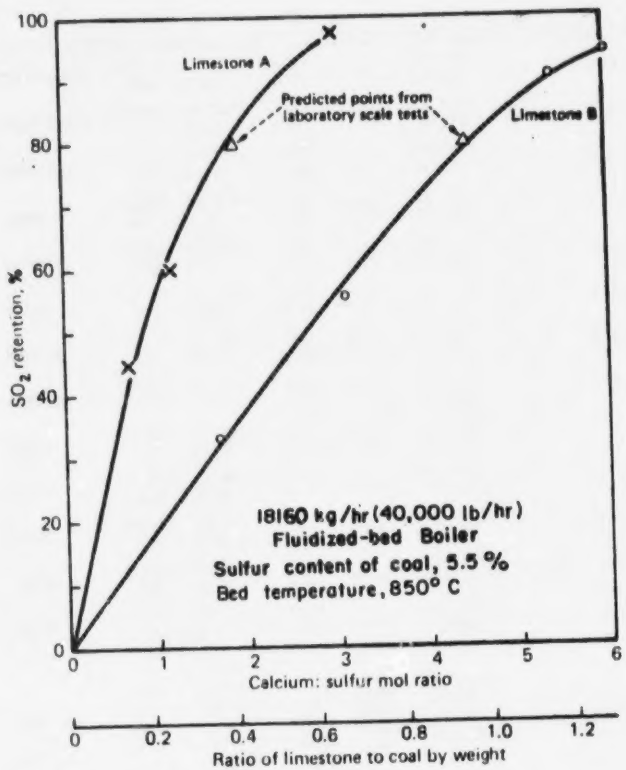
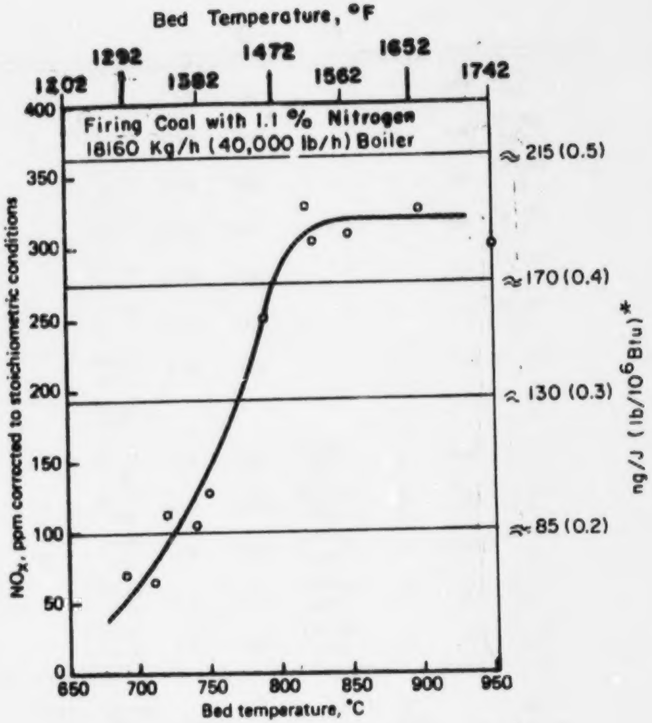


Figure 57. Results of SO_2 emission testing at Renfrew, Scotland FBC boiler reported by B&W, Ltd. (Courtesy of Babcock Contractors, Inc.)



*ESTIMATED

Figure 58. Results of NO_x emission testing at Renfrew, Scotland FBC boiler reported by B&W, Ltd. (Courtesy of Babcock Contractors, Inc.)

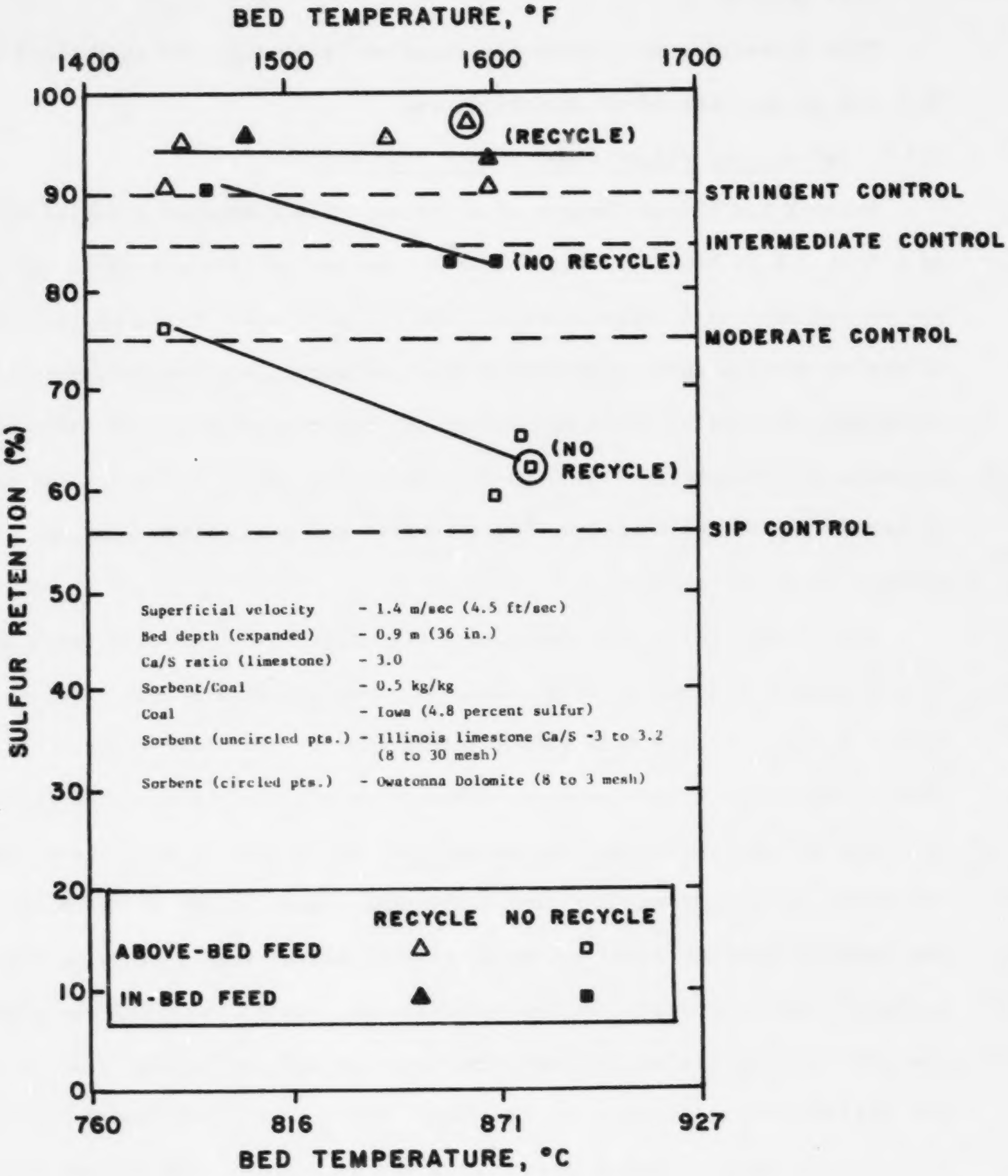


Figure 59. Sulfur retention data in FluiDyne's 0.46 × 0.46 m (1.5 ft × 1.5 ft) FBC unit.[11]

7.3 TEST METHODS

This subsection delineates the sampling technology and analytical procedures followed by the individual investigators.

7.3.1 Babcock and Wilcox (B&W) 6 ft × 6 ft Unit[17,18,19]

Babcock and Wilcox Company of Alliance, Ohio conducted a series of tests in a 6 ft × 6 ft fluidized-bed combustion boiler in 1978 and 1979. The project was established as a cooperative B&W and Electric Power Research Institute (EPRI) effort to develop sufficient design data and accumulate convincing operating experience in a pilot scale FBC boiler to justify demonstration and commercialization of atmospheric fluidized-bed combustion (AFBC) boilers. The data collected in the tests include SO_2 emissions and particulate loadings at the cyclone inlet and outlet.

The Furnace Outlet Gas Sampling Probe (Figure 60), through which the SO_2 data reported in Table 81 were collected, consists of a sheath (cooling jacket) around a single-center tube with a quartz liner. The liner extends beyond the rear of the metal sheath where it connects to a cyclone oven. The cyclone oven is a heated box containing a glass cyclone, catch bottle, and filter assembly. The probe is operational any time combustion occurs in the 6 ft × 6 ft unit. Gas samples from the probe are drawn through heated sample lines to the Beckman® analyzer system in the control room. An NO_x analyzer was added to the gas sampling system for more comprehensive testing during 1979. Details of the analyzing systems were not reported. The Cyclone Inlet and Cyclone Outlet Particulate (Dust) Sampling System consists of a probe, electropneumatic control valve, transducer, condenser, vacuum gauge, gas meter, mounting flange, Bug-O® drive, and vacuum pump (Figure 61). Traversing was performed automatically using the Bug-O® drive unit. Figure 62 illustrates the probe and its internal

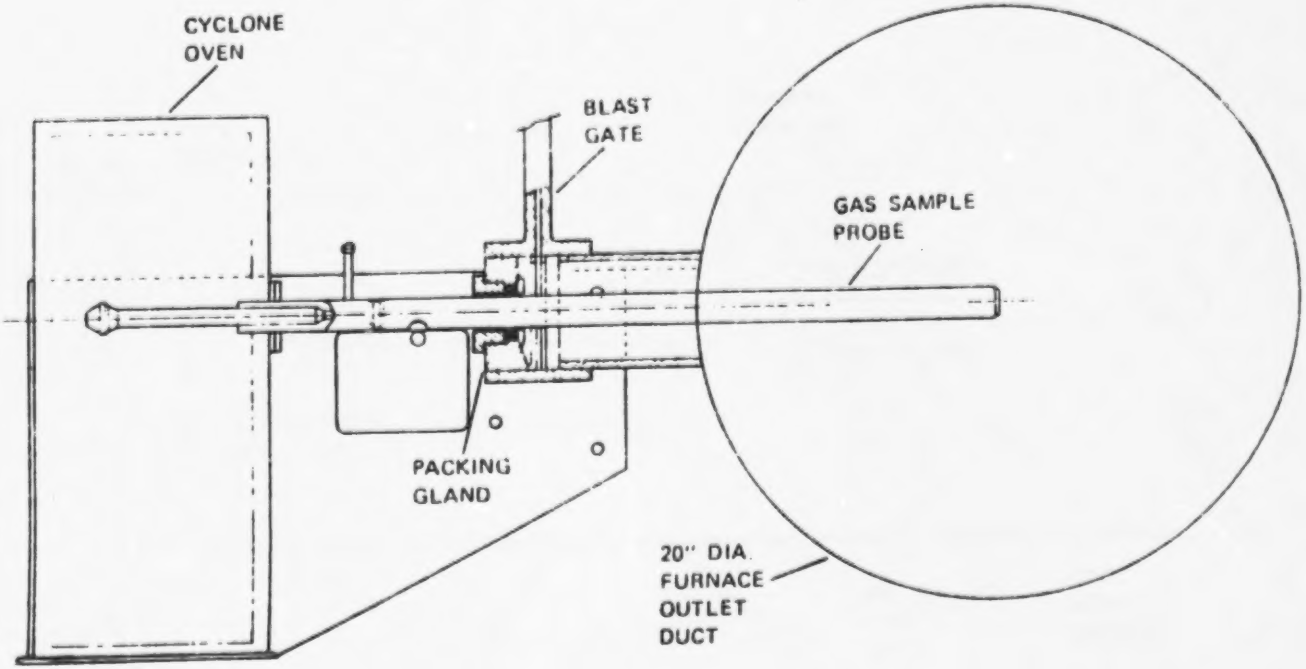

CYCLONE
OVEN
BLAST
GATE
GAS SAMPLE
PROBE
PACKING
GLAND
20" DIA.
FURNACE
OUTLET
DUCT

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Figure 60. Furnace outlet gas sampling system for EPRI/B&W 6 ft × 6 ft unit. **(Reproduced with permission of EPRI.)**

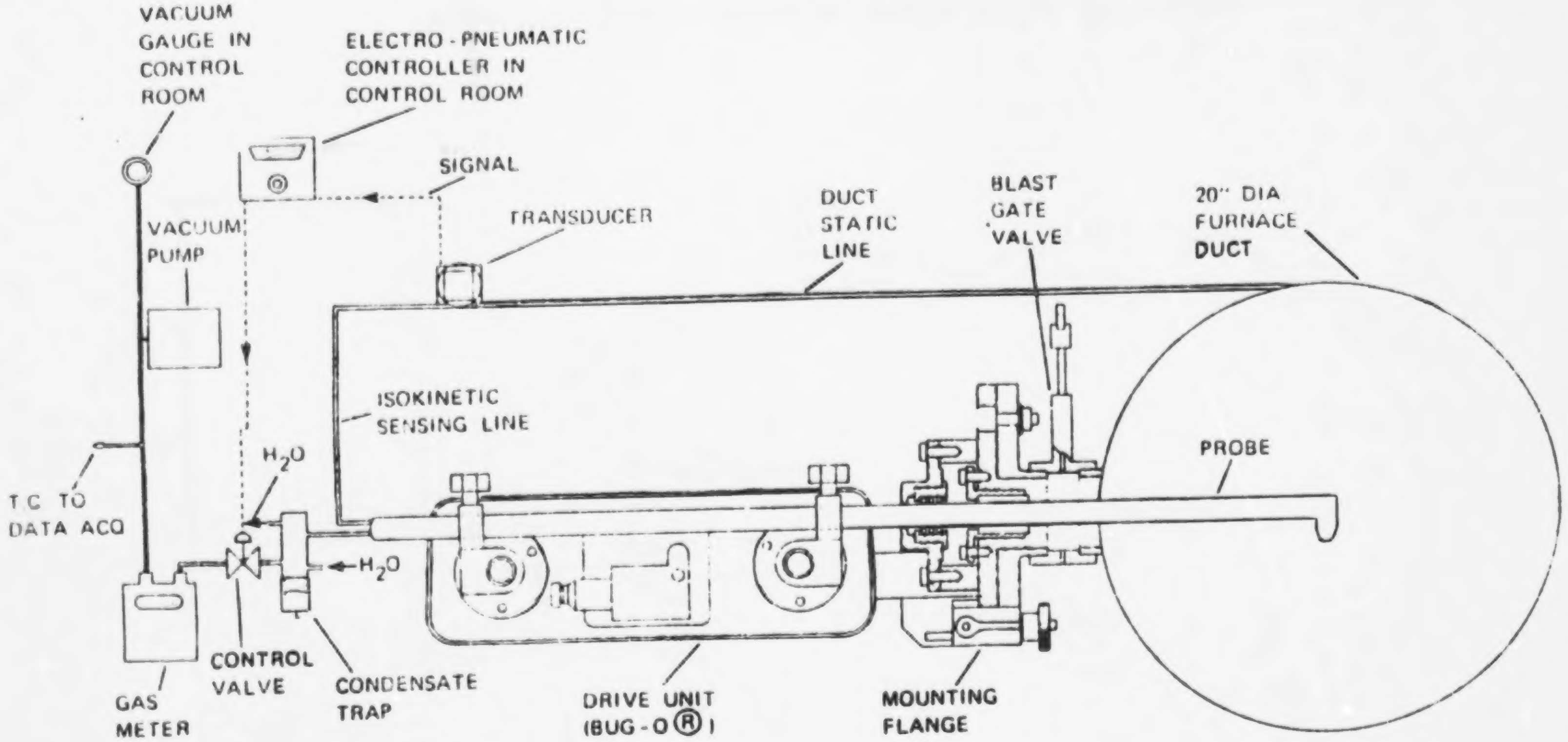
VACUUM GAUGE IN CONTROL ROOM
ELECTRO-PNEUMATIC CONTROLLER IN CONTROL ROOM
SIGNAL
TRANSDUCER
DUCT STATIC LINE
BLAST GATE VALVE
20" DIA FURNACE DUCT
VACUUM PUMP
ISOKINETIC SENSING LINE
PROBE
H_2O
T.C TO DATA ACQ
H_2O
GAS METER
CONTROL VALVE
CONDENSATE TRAP
DRIVE UNIT (BUG-O®)
MOUNTING FLANGE

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Figure 61. Arrangement of cyclone inlet and outlet dust sampling equipment for EPRI/B&W 6 ft × 6 ft unit. (Reproduced with permission of EPRI.)

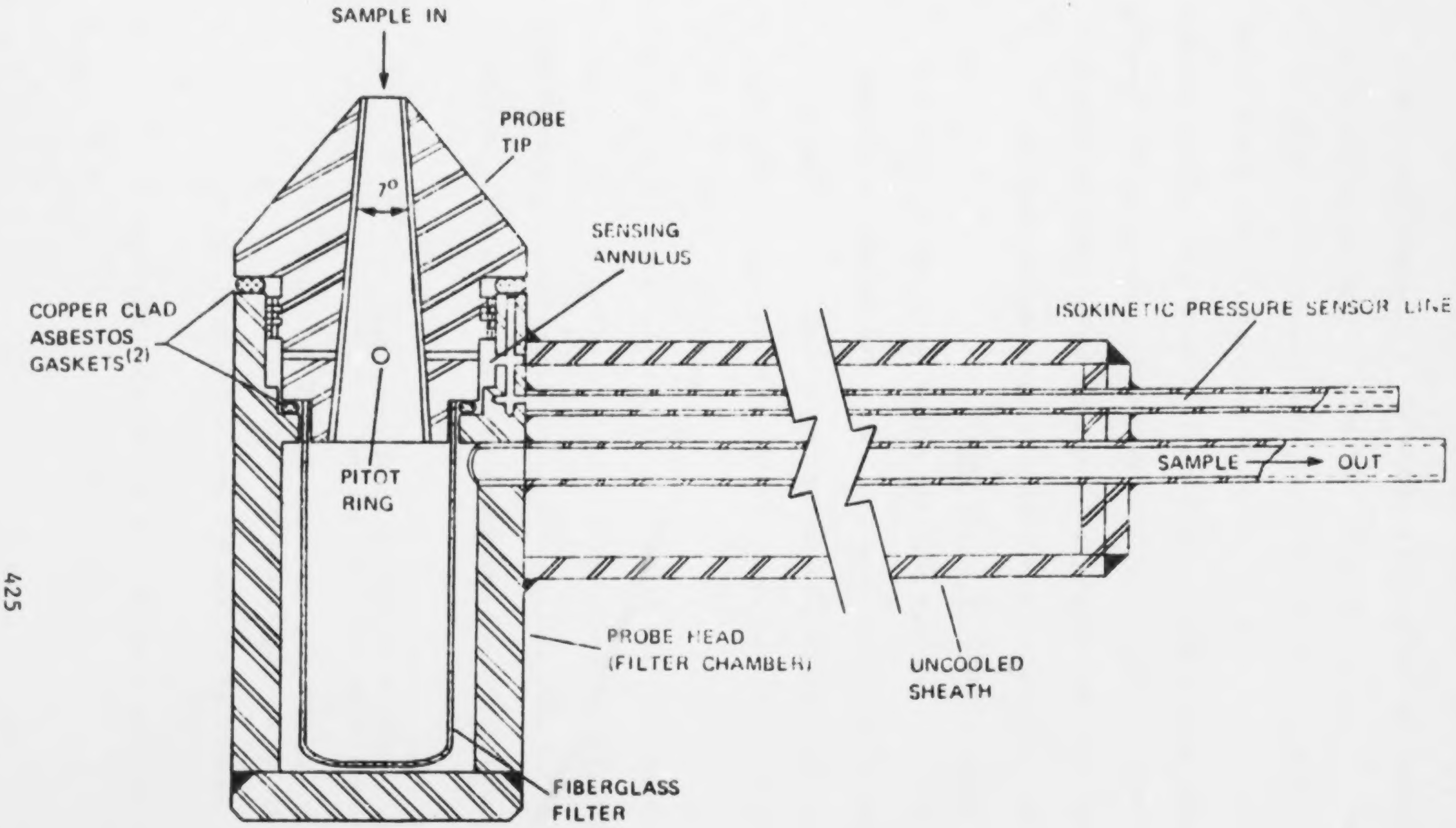
SAMPLE IN
PROBE TIP
7°
SENSING ANNULUS
COPPER CLAD ASBESTOS GASKETS(2)
ISOKINETIC PRESSURE SENSOR LINE
PITOT RING
SAMPLE → OUT
PROBE HEAD (FILTER CHAMBER)
UNCOOLED SHEATH
FIBERGLASS FILTER

Figure 62. Cyclone inlet and outlet dust sampling probe for EPRI/B&W 6 ft × 6 ft unit. **(Reproduced with permission of EPRI.)**

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fiberglass filter. It is not clear from available information whether the probe orientation of this unit or any of the units discussed meets the requirements of the EPA Reference Methods. The main body of this probe is an uncooled sheath containing two tubes. The large tube connects the probe to the condenser while the other connects it to the transducer. This probe is an isokinetic type, based on null balance techniques. During null balance isokinetic sampling, an attempt is made to equalize the static pressure in the sampling duct and in the probe tip. Maintaining this balance during sampling insures that a representative (isokinetic) dust sample is taken during testing.

7.3.2 Babcock and Wilcox (B&W) 3 ft × 3 ft Unit[20]

Babcock and Wilcox has reported results of testing on their 3 ft × 3 ft fluidized-bed combustion facility during late 1976. The purpose of the B&W testing was to assess the effect of sorbent particle size on SO_2 absorption. The data acquired in the tests covers the three major pollutants - SO_2, NO_x, and particulates.

Emissions were sampled at the inlet of a wet scrubber attached to the unit. Test duration was normally between 6 and 8 hours, and emissions data were acquired after the unit had been equilibrated at the desired operating condition.

The following sampling and measurement procedures were normally carried out:

- Coal feed, sorbent feed, bed material and hopper ash were each sampled at the start and end of each test;
- Flue gas at the scrubber inlet was sampled and analyzed for SO_2, O_2, CO and NO_x throughout each test;
- Spot measurements of CO_2 and H_2S were made at the scrubber inlet; and
- Dust loadings were measured over a five-point traverse at the scrubber inlet.

The following methods of gas analysis for SO_2 and NO_x were used:

- SO_2 - DuPont Model 411 (light absorption in uv range)
  - Barton Model 256 (continuous titration of SO_2/H_2S with bromine)
  - Reich wet chemical spot check (titration of SO_2 with potassium iodate)[21]
- NO_x - Teco Model 10A (chemiluminescence from reaction with ozone).

Sampling of flue gas at the scrubber inlet employed the sampling probe shown schematically in Figure 63. The probe was lined with a 7-mm I.D. quartz tube. The suction rate through the probe was normally 6 to 7 l/min. Water cooling was not used in all tests. The oven temperature was maintained near 250°F, and the impinger-exit temperature was maintained below room temperature. Figure 72 shows the overall gas analysis system applied at the scrubber inlet.

The DuPont SO_2 measurement was supplemented during part of each test by measurements in the Barton instrument. Comparisons of the different methods of measurement of SO_2 at the scrubber inlet were also made. The two methods of SO_2 measurement generally agreed within ±12 percent. The scrubber-inlet SO_2 measurements in Table 82 are from the DuPont instrument.

Dust loading was measured during each test at the scrubber inlet for 1 hour. A five-point equal-area traverse was made at the scrubber inlet duct. The probe used to measure dust loading at the scrubber inlet is shown in Figure 65. The sample gas rate was adjusted to give an isokinetic inlet velocity.

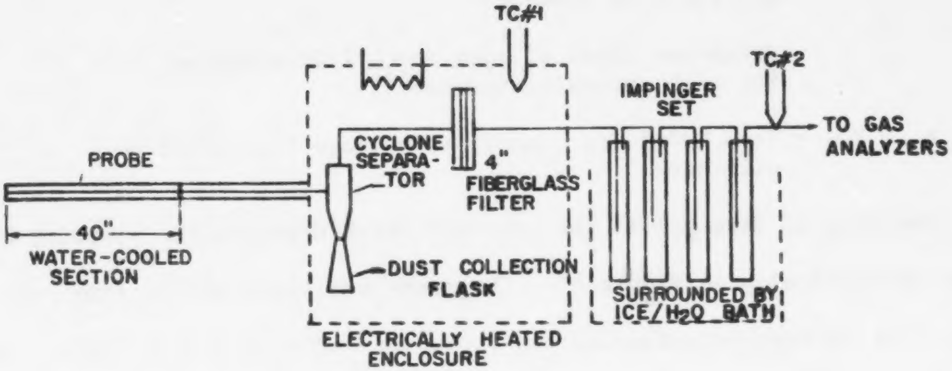
TC#1
TC#2
IMPINGER SET
TO GAS ANALYZERS
PROBE
CYCLONE SEPARATOR
4" FIBERGLASS FILTER
40"
WATER-COOLED SECTION
DUST COLLECTION FLASK
SURROUNDED BY ICE/H2O BATH
ELECTRICALLY HEATED ENCLOSURE

Figure 63. Gas sampling system employed by B&W at wet scrubber inlet. (Reproduced with permission of EPRI.)

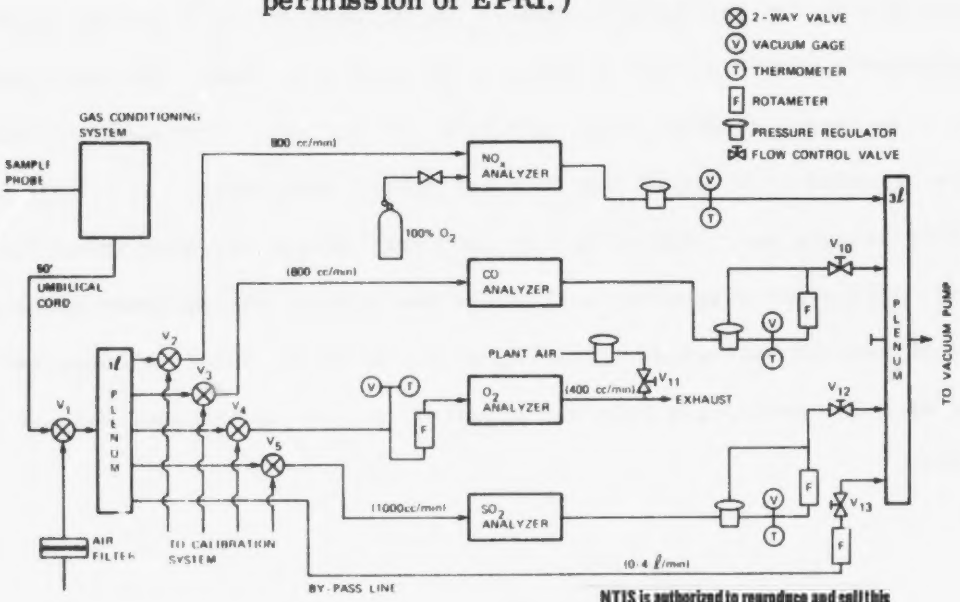
2-WAY VALVE
VACUUM GAGE
THERMOMETER
ROTAMETER
PRESSURE REGULATOR
FLOW CONTROL VALVE
GAS CONDITIONING SYSTEM
SAMPLE PROBE
800 cc/min)
NOx ANALYZER
100% O2
3l
V10
(800 cc/min)
CO ANALYZER
50' UMBILICAL CORD
PLENUM
TO VACUUM PUMP
PLANT AIR
V2
V3
V1
1l
PLENUM
V4
V5
V11
(400 cc/min)
EXHAUST
O2 ANALYZER
V12
F
V13
(1000cc/min)
SO2 ANALYZER
AIR FILTER
TO CALIBRATION SYSTEM
(0.4 l/min)
BY-PASS LINE

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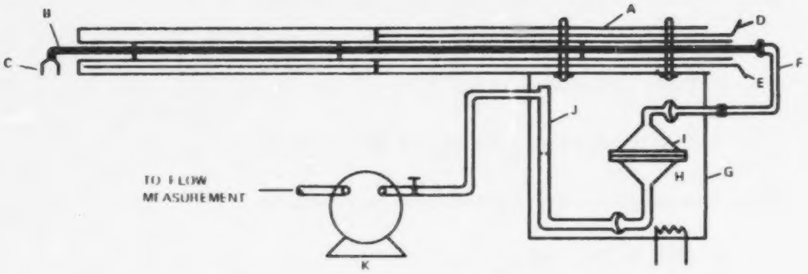
Figure 64. Overview of gas sampling and analysis system employed by B&W at wet scrubber inlet. (Reproduced with permission of EPRI.)



A	WATER JACKET PROBE – 1-1/2" O.D. – 60" LENGTH STAINLESS STEEL 9/16" I.D. – 62" LENGTH
B	GAS INLET LINE – 3/8" O.D. S.S. TUBING – 68" LONG
C	INLET NOZZLE – 3/4" O.D. S.S. TUBING 1/2" LENGTH – 0.652" I.D.
D	PROBE H_2O OUTLET – 1/8" S.S. TUBING
E	PROBE H_2O INLET – 1/8" S.S. TUBING TO WITHIN 1/4" OF END
F	CONNECTOR LINE – 5/16" I.D. TYGON TUBING
G	HEATED FILTER BOX – SCIENTIFIC GLASS & INST. CO. (250°F) #5000 – 28G4
H	4" DIA. FILTER HOLDER – SCIENTIFIC GLASS & INST. CO. #1010-4-4
I	112 mm GLASS FILTER – S.G.I. CO. – #1002-FG4 (0.3 MICRON)
J	SI. EDGE PIPE ORIFICE – 1/8" ORIFICE IN 1/2" SCH. 40 PIPE
K	VACUUM PUMP

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Figure 65. Particulate sampling probe used in B&W investigations. (Reproduced with permission of EPRI.)

7.3.3 National Coal Board - 3 ft × 1.5 ft Unit[22,23]

Flue gas samples were taken in the stack downstream of the secondary cyclone. Three methods for sulfur dioxide and two methods for nitrogen oxides analysis provided information concerning pollutant concentrations in the flue gas from the CRE unit. The methods were as follows:

- SO_2 - Continuous online Hartman-Braun infrared analyzer.
  - The iodine method.[24]
  - The hydrogen peroxide method.[25]
- NO_x - A modified Saltzman's method.[26]
  - The BCURA NO_x box.[27]

The iodine method was the standard method used to determine SO_2 concentration. Flue gas was bubbled through an iodine solution and SO_2 concentration was determined colorimetrically. Using the hydrogen peroxide method, flue gas was bubbled through a solution of hydrogen peroxide and the sulfate produced was determined gravimetrically by precipitation as barium sulfate. The Hartman-Braun analyzer was run continuously and all results were compared periodically.

To determine NOx, a modified Saltzman's method was used by drawing a sample of SO_2 free gas into an evacuated 500 ml sample bottle containing 40 ml of Saltzman's reagent. At 30-min intervals, solution was withdrawn and fresh reagent was added. This was repeated until the color developed by the solution was negligible. All of the solution was bulked and the intensity of the color was measured using a spectrophotometer.

In the BCURA NO_x box the SO_2 free gas is first passed through an oxidizer in which any NO present is converted to NO_2. The gas is then passed through a cell containing a platinum gauze electrode moistened by a wick dipping into an electrolyte solution in which an active carbon electrode was immersed.

A microammeter was used to measure the current through the external circuit through the electrodes, which varied as a function of NO_2 concentration.

The data reported are the result of testing Pittsburgh and Welbeck coals with limestone 18, dolomite 1337, and U.K. limestone as sorbents. Since the main test objectives were correlation of parametric effects on emissions with data obtained in a smaller unit rather than demonstration of operating reliability, no long-term testing was attempted. Typical test duration at steady-state at a specific set of operating conditions ranged from 2 to 4 hours. This did not include startup or condition changes.

7.3.4 Pope, Evans, and Robbins[28] [29]

The emission test data reported by PER and presented in Tables 84 and 85 were compiled from experiments conducted between 1967 and 1975 Gas samples were withdrawn from the FBM at the gas passage around the steam drum through a 7.6 cm (3 in.) diameter welded pipe. A schematic diagram of the sampling system is shown in Figure 66.

Emissions of SO_2 and NO_x were monitored continuously by infrared (Beckman Model 215) analysis and periodically checked using methods similar to EPA Reference Methods 6 and 7.

Particulate emissions were monitored using an isokinetic probe system at one point. The sampling location was downstream of the multicone collector and prior to the ID fan (see Figure 71).

The test procedures for the FBM investigations involved igniting the bed and stabilizing the combustion at the desired bed temperature until steady-state conditions prevailed. Steady-state was assumed when the Bailey Meter used for O_2 measurement and the SO_2 IR analyzer indicated constant values of oxygen and sulfur dioxide in the flue gas. At steady-state the sorbent feed

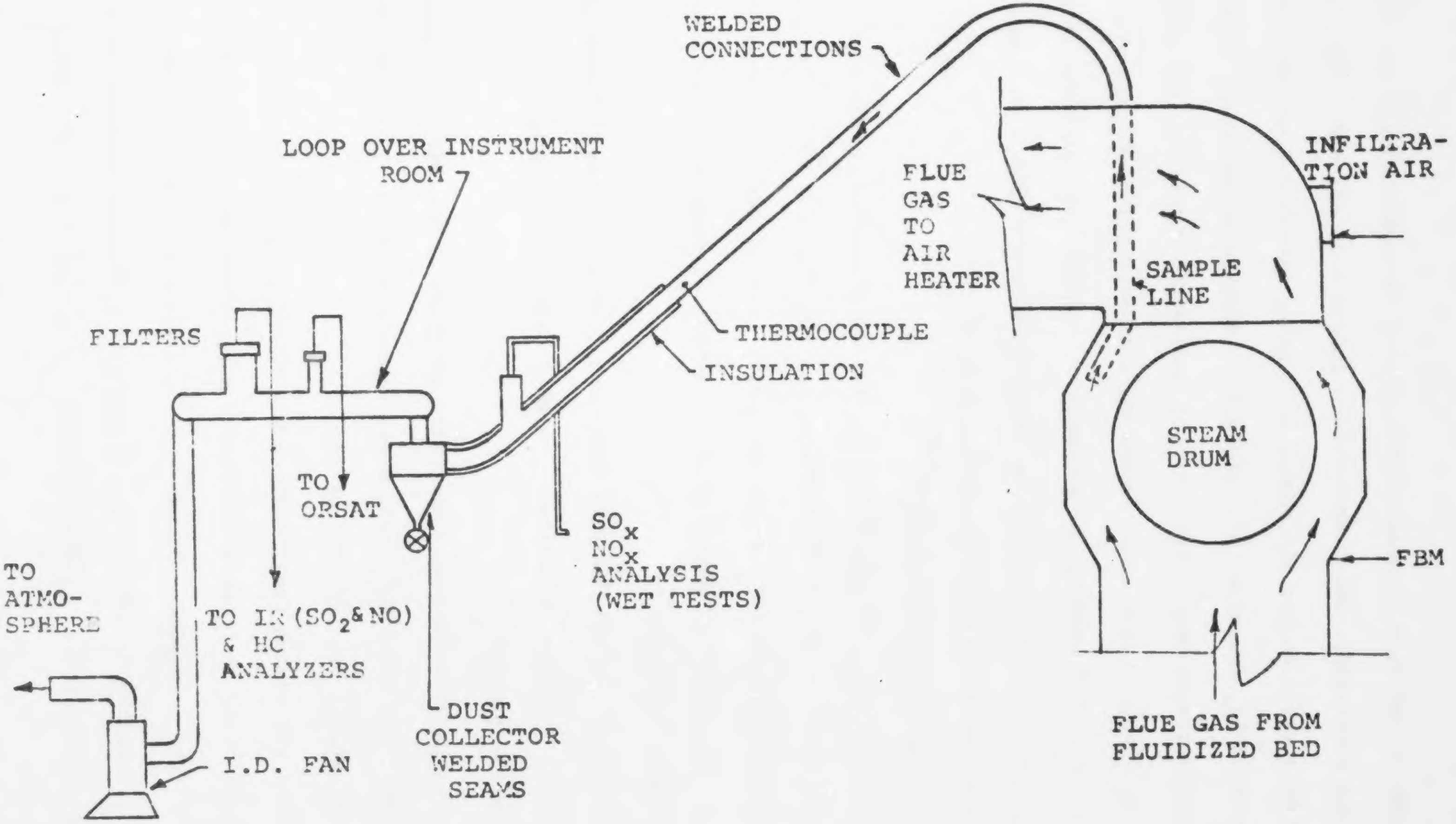
WELDED CONNECTIONS
LOOP OVER INSTRUMENT ROOM
FLUE GAS TO AIR HEATER
INFILTRATION AIR
SAMPLE LINE
FILTERS
THERMOCOUPLE
INSULATION
STEAM DRUM
TO ORSAT
SO$_x$ NO$_x$ ANALYSIS (WET TESTS)
FBM
TO ATMOSPHERE
TO IR (SO$_2$ & NO) & HC ANALYZERS
I.D. FAN
DUST COLLECTOR WELDED SEAMS
FLUE GAS FROM FLUIDIZED BED

Figure 66. Schematic diagram of gas sampling system used by PER during FBM experiments.

was initiated or some other operating condition varied and the effect on emission observed. A period of 30 min, at least, was allowed for a new steady-state condition after an operating condition change. Each run lasted from 2 to 6 hours.

7.3.5 FluiDyne[30,31]

Emissions testing equipment used to monitor the 1.5 ft × 1.5 ft unit and the 3.3 ft × 5.3 ft vertical slice combustor, included the following:

- Gas Composition Measurement Instrumentation
    - Beckman Model 864 NDIR CO_2 Analyzer
    - Beckman Model 865 NDIR SO_2 Analyzer
    - Beckman Model 742 O_2 Analyzer
    - Fisher Orsat (CO measurement)
    - DuPont Model 411 SO_2/NO_x Analyzer
- Flue gas Particulate Measurement Instrumentation
    - Water cooled sampling probe with alumina thimble holder
    - Blue M Globar (15 kw) furnace and analytical balance

During sampling of the vertical slice combustor, SO_2 measurements using the Beckman Model 865 were checked by including the DuPont Model 411 in the flue gas sampling system. It is noted that the Beckman unit consistently indicated flue gas SO_2 concentrations higher than actual (based on wet chemical tests and readings from the DuPont instrument), so that reported sulfur retention levels should be conservative.

7.3.6 National Coal Board - 6-in. Diameter Unit[32]

Gas samples were withdrawn at a point about 2 ft after each secondary cyclone, as appropriate, (see Figure 85), and bubbled through iodine or H_2O_2 solution for determination of SO_2. Samples were also taken for analysis of O_2, CO, CO_2 and CH_4 by gas chromatograph.

Each test was carried out as a 1-day (16-hr) run comprising plant startup, approach to equilibrium, a 6-hr mass balance and shutdown.

7.3.7 Argonne National Laboratory (ANL)[33,34,35]

The data presented in Table 91 was obtained from the ANL 6-in. diameter atmospheric pressure fluidized-bed combustion unit.

The sampling methods used for the system follow. A continuous stream of approximately one-twentieth of the total flue gas (0.24 l/sec) was withdrawn through a 1.3 cm (0.5 in.) diameter stainless steel sample probe from the upper portion of the bench-scale unit. The gas was dried by passage through a water condenser and refrigerator. Continuous analysis of NO and SO_2 was carried out using Beckman 315A infrared analyzers. Figure 67 is a general schematic of the system.

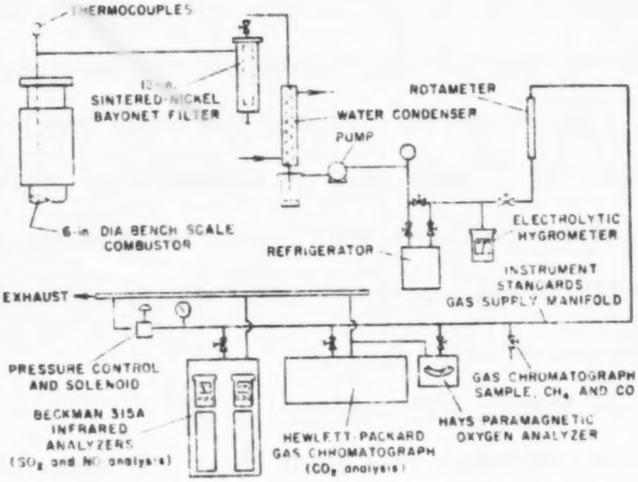
THERMOCOUPLES
12-in. SINTERED-NICKEL BAYONET FILTER
ROTAMETER
WATER CONDENSER
PUMP
6-in DIA BENCH SCALE COMBUSTOR
ELECTROLYTIC HYGROMETER
REFRIGERATOR
INSTRUMENT STANDARDS GAS SUPPLY MANIFOLD
EXHAUST
PRESSURE CONTROL AND SOLENOID
GAS CHROMATOGRAPH SAMPLE, CH$_4$ AND CO
BECKMAN 315A INFRARED ANALYZERS (SO$_2$ and NO analysis)
HAYS PARAMAGNETIC OXYGEN ANALYZER
HEWLETT-PACKARD GAS CHROMATOGRAPH (CO$_2$ analysis)

Figure 67. ANL gas sampling and analysis system.

7.3.8 Babcock and Wilcox, Ltd.[36]

Limited information is available on testing procedures at the B&W, Ltd. unit located in Renfrew, Scotland. The currently available publication indicates only that NO_x was measured using a chemiluminescence monitor.

7.4 DESCRIPTION OF TEST FACILITIES

7.4.1 Babcock and Wilcox (B&W) - 6 ft × 6 ft Unit[37 38 39]

The B&W 6 ft × 6 ft AFBC unit has four feed points at the spacing of one feedpoint per 9 ft^2 of bed area with allowance for operating with fewer feed points.* The unit was designed to produce steam for heating the Alliance Research Center (ARC) B&W's research facility. Condensing the steam and recycling treated water back to the unit provides operational cost savings.

Once the overall bed size and steam producing capabilities were defined, the other basic design parameters listed below were established.

6 ft × 6 ft Design Parameters (Nominal)

Bed Area	6 ft × 6 ft
Superfical Velocity	8 fps
Coal Feed Rate	1880 lb/hr
Heat Rate	$\approx 7\ MW_t$
Saturated Steam Production	10,000 lb/hr at 150 psig
Superheated Steam Production	2,000 lb/hr at 1000^oF
Bed Operating Temperature	$\sim 1600^oF$

Figure 68 identifies the major components of the facility. Coal and limestone are conveyed to the top of the Boiler Room where they are crushed, then transported either directly to two separate bunkers or through an intermediate screening operation. Coal and limestone from the bunkers are fed through separate weigh feeders into a common transport line. The feed solids

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*This unit may be modified to use fewer feedpoints, during 1979.

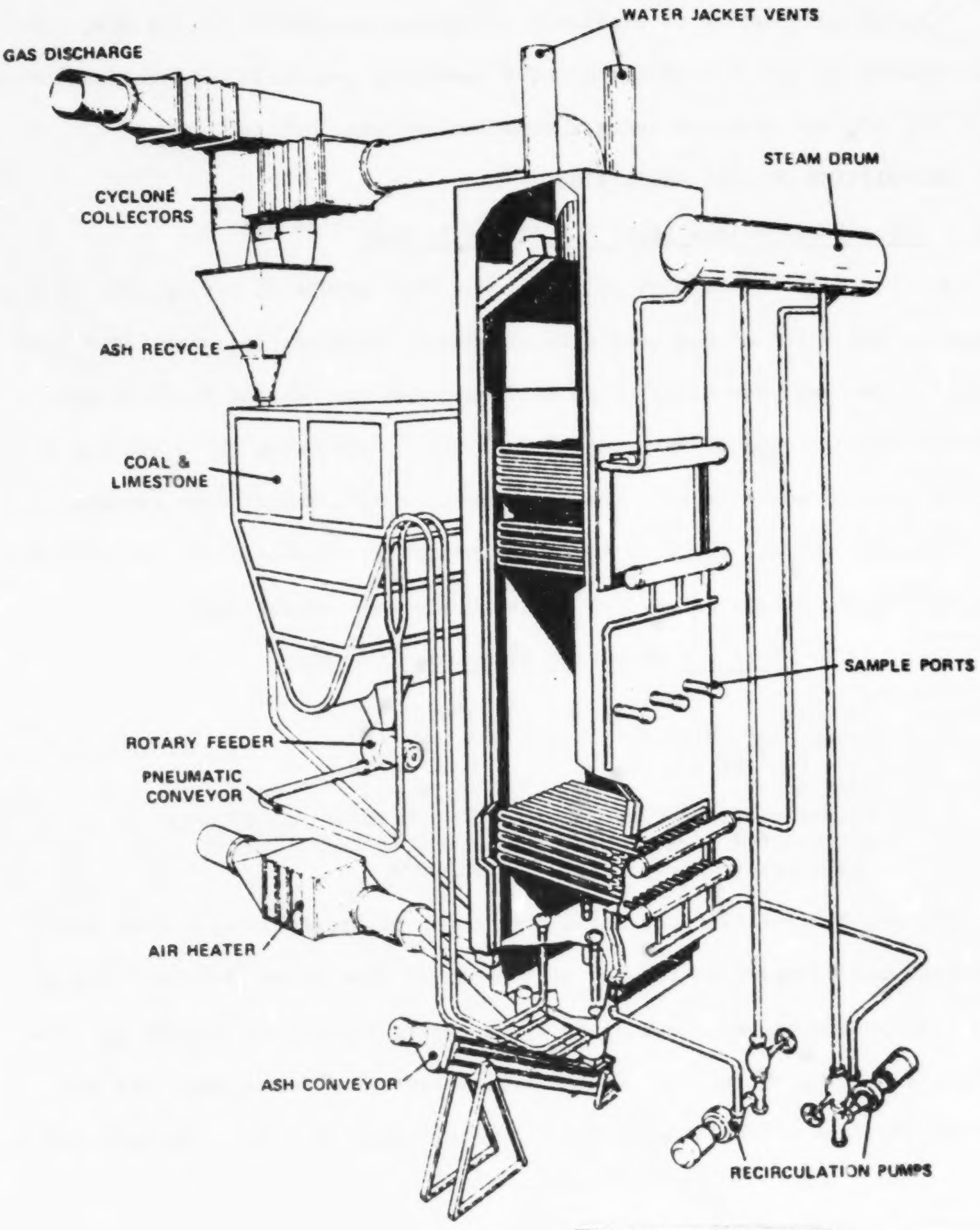
WATER JACKET VENTS
GAS DISCHARGE
STEAM DRUM
CYCLONE COLLECTORS
ASH RECYCLE
COAL & LIMESTONE
SAMPLE PORTS
ROTARY FEEDER
PNEUMATIC CONVEYOR
AIR HEATER
ASH CONVEYOR
RECIRCULATION PUMPS

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Figure 68. Fluidized-bed combustion development facility (B&W). (Reproduced with permission of EPRI.)

are picked up by transport air and carried to a splitter where they are separated into four equal feed streams. These pass up through the windbox and the distributor plate into the combustion zone of the fluidized-bed boiler.

Forced draft air to the fluidized-bed is supplied by a Spencer turbine centrifugal blower capable of delivering 6,000 cfm at a 60-in. water gauge head. The combustion air supplied by this fan first passes through a steam preheater and then through a direct-fired preheater before it diverges into the four separate ducts entering the windbox. Each of these ducts has a separate damper and venturi flowmeter for control and measurement.

The initial distributor plate was made of woven Ni-Chrome wire that had been calendered to obtain a specified pressure drop at a design flow rate per square foot of bed area (10 in. water pressure drop at 8 ft/sec). Because of warpage and pluggage problems, the woven wire was replaced with a perforated distributor plate. This plate is type 316 stainless steel having 0.0938 in. holes on 0.587 in. square pitch. The distributor plate and windbox are designed as a unit that can be lowered from 20 in. below (initial position) to 40 in. below the immersed tube bank.

The main furnace structure of the fluidized-bed test facility consists of an atmospheric pressure water wall with fireside refractory lining.

The immersed tube bank consists of a serpentine arrangement of 11, 1-1/2 in. O.D. tubes on a 5 in. triangular pitch.

One tube is used as a superheater. The balance of the tube bank consists of steam generating tubes which will produce 150 psig saturated steam.

A freeboard of 18 ft is located between the immersed tube bank and the convective tube bank at the top of the furnace. This height was chosen so that the larger particles thrown out of the bed would return to the bed; i.e, particles with a terminal settling velocity greater than the fluidizing velocity would fall back.

The convective tube bank at the top of the furnace serves two purposes. First, it cools the flue gas before it exits the furnace and enters the cyclone dust collectors. Second, it produces additional saturated steam for heating the Alliance Research Center. Space in the center of this tube bank has been allotted for a sootblower, if one is found necessary.

Four cyclone separators are mounted at the furnace exit to collect particulates escaping the furnace. Dampers on each of the cyclones can be closed to maintain reasonable entering velocities and, by so doing, improve collection efficiencies. Material collected by the cyclones can be recycled to the bed or removed from the unit by the ash-handling system. Material to be recycled is fed from the cyclone hoppers through a water-cooled conveyor. After passing over an inline impact flowmeter, the material passes through a downcomer to the transport air line in the coal and limestone feed system. The recycle system as initially designed is capable of recycling only about one-quarter of the carryover back to the combustor. Testing reported in this section was conducted with this recycle capability. The system is currently being modified to enable full recycle.

The flue gas exiting the cyclones passes through a large venturi flowmeter and then is cooled before entering the induced draft fan which carries it out the stack.

The boiling water circuit consists of a split steam drum and two recirculation pumps which feed the immersed and convective tube banks. Separate makeup and blowdown systems are also provided.

The spent bed removal system consists of five drain pipes which extend from the bed through the windbox and the distributor plate to the basement. Each pipe has a separate shutoff valve controlled by an air cylinder. Initially, only the center pipe will be used to remove material from the bed. During an upset condition all of the pipes can be opened to rapidly drain the bed of solids. The rate of bed removal is controlled by the pressure drop across the bed. This system can be easily modified so that the control of solids removal is set either by bed temperature, the input limestone and coal feed rates, and/or a time sequence.

7.4.2 Babcock and Wilcox 3 ft × 3 ft Unit[40]

The 3 ft × 3 ft unit is a vertical furnace enclosed by an atmospheric pressure water-jacket. Fluidizing air is supplied to the furnace by a 3,500 rpm fan rated for 4.25 m^3/sec (9,000 cfm) at a pressure of 13.7 kPa (55 in. of water). Coal and limestone are generally crushed, screened, and sized prior to charging. Coal feed rate can be varied from 90.9 to 1,818 kg/hr (200 to 4,000 lb/hr), and limestone feed rate can be varied from 45.4 to 909 kg/hr (100 to 2,000 lb/hr). Coal and limestone are added to the boiler as a mixture. A boiler tube bank is positioned in the bed consisting of 8.9 square meters (96 ft^2) of cooling surface. The tubes are cooled by recirculating cooled water at approximately 1,172 kPa (170 psig). Primary flue gas particulate removal was provided by a larger water-jacketed cavity in the flue. During testing, fly ash recirculation was not practiced. The freeboard in this unit is low and primary collection efficiency is poor so that particulate carryover is high. A schematic diagram of the Babcock and Wilcox 3 ft × 3 ft FBC appears in Figure 69.

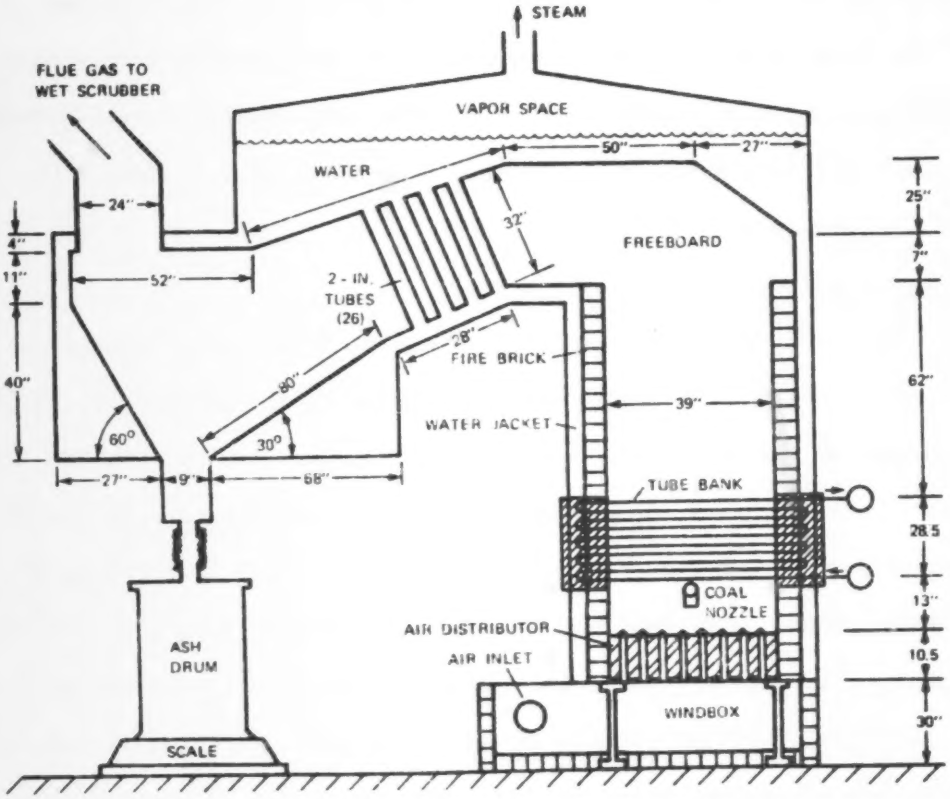
STEAM
FLUE GAS TO WET SCRUBBER
VAPOR SPACE
WATER
50"
27"
24"
4"
11"
52"
40"
60°
27"
9"
68"
30°
80"
2 - IN. TUBES (26)
32"
28"
FREEBOARD
FIRE BRICK
WATER JACKET
39"
25"
7"
62"
TUBE BANK
28.5
13"
COAL NOZZLE
AIR DISTRIBUTOR
AIR INLET
10.5
WINDBOX
30"
ASH DRUM
SCALE

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Figure 69. Schematic diagram of B&W 3 ft × 3 ft test unit. (Reproduced with permission of EPRI.)

7.4.3 National Coal Board 3 ft × 1.5 ft Unit[41,42]

The Coal Research Establishment (CRE) unit has an internal cross-section of 0.9 × 0.46 meters (3 ft × 1.5 ft). The height from the air distributor to the gas off-take was 4.6 meters (15 ft). Coal and limestone were pneumatically fed in adjacent lines to the center of the bed. Off-gases pass through primary and secondary cyclones and then to the stack. The primary recycle capability of the unit was not utilized during the tests for which data is reported. Fourteen water-cooled tubes of 5 cm (2 in.) inside diameter are included in the bed. Coal feed rate is variable between 34 to 136 kg/hr (75 to 300 lb/hr). A schematic diagram of the boiler is shown in Figure 70.

7.4.4 Pope, Evans, and Robbins FBM Unit[43]

The PER-FBM was intended to represent one-half of a multicell FBC package boiler. The 1.5 ft × 6 ft rectangular bed was surrounded by vertical water tubes and an overhead drum. There were no boiler tubes located through the bed. Flue gas passed around the steam drum. Freeboard in the boiler was short, the total distance from grid to bottom of steam drum was only 1.6 meters (5 ft, 4 in.). The combustion space was 1.5 m^3 (53 ft^3) with a projected heating surface of 7.4 m^2 (80 ft^2). Boiler capacity is 2,270 kg/hr (5,000 lb/hr) steam excluding convection heat transfer and 3,180 kg/hr (7,000 lb/hr) including convection heat transfer. Pressure rating is 3,070 kPa (300 psi) design and 1,380 kPa (200 psi) normal operation. Coal feed varies between 300 to 400 kg/hr (700 to 900 lb/hr). A multicone dust collector and hopper is included which contains 12, 25 cm (10 in.) diameter centrifugal collector units, a rotary feeder for fly ash reinjection and valve for fly ash removal. Fly ash reinjection was possible as an option, and was employed in a few, but not the bulk of the tests summarized in Table 84. A schematic diagram of the FBM appears in Figure 71.

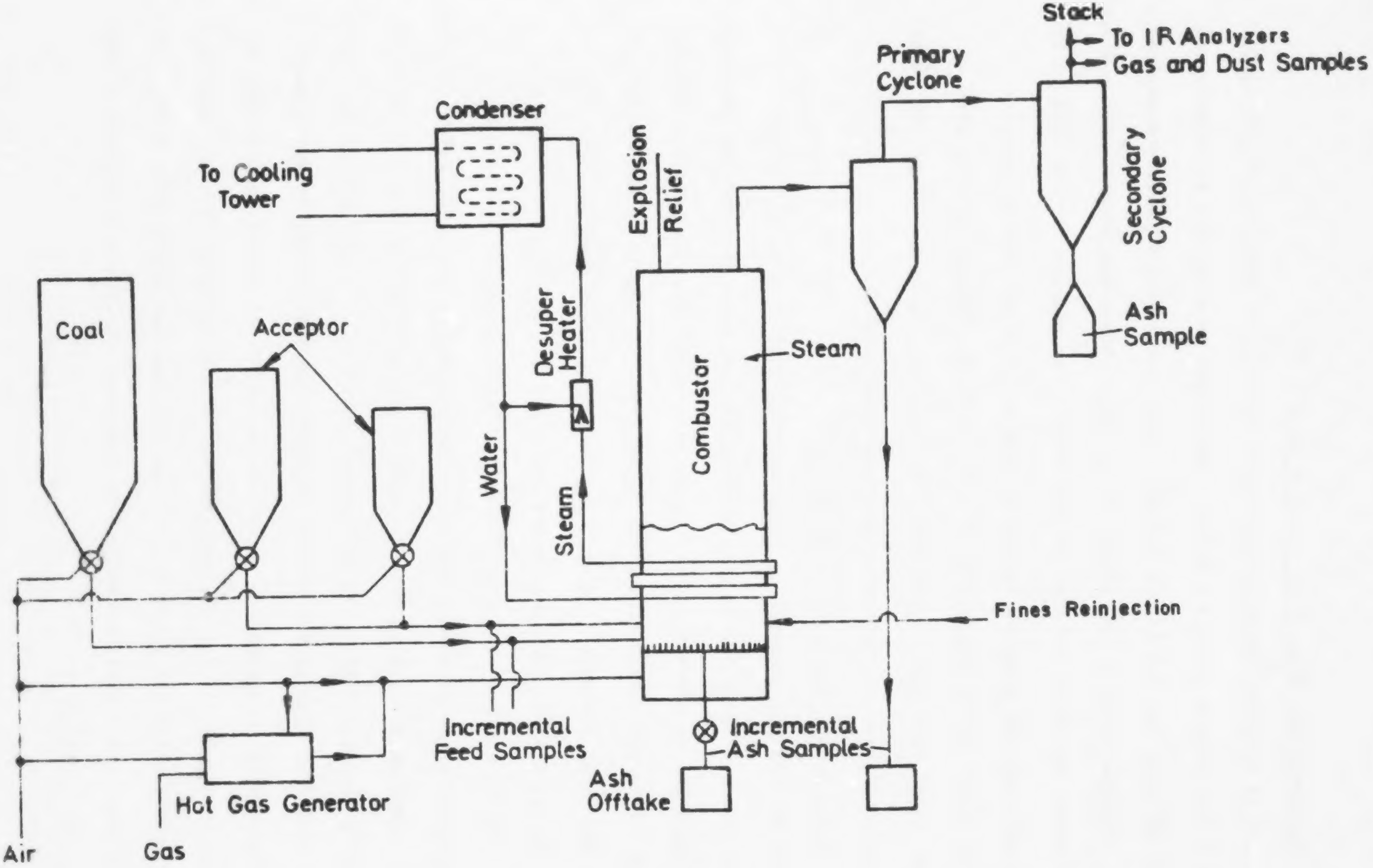
Stack
To I R Analyzers
Gas and Dust Samples
Primary Cyclone
Condenser
To Cooling Tower
Explosion Relief
Secondary Cyclone
Coal
Acceptor
Desuper Heater
Combustor
Steam
Ash Sample
Water
Steam
Fines Reinjection
Incremental Feed Samples
Incremental Ash Samples
Ash Offtake
Hot Gas Generator
Air
Gas

Figure 70. Schematic diagram of CRE 18 in. × 72 in. FBC facility tested by NCB.

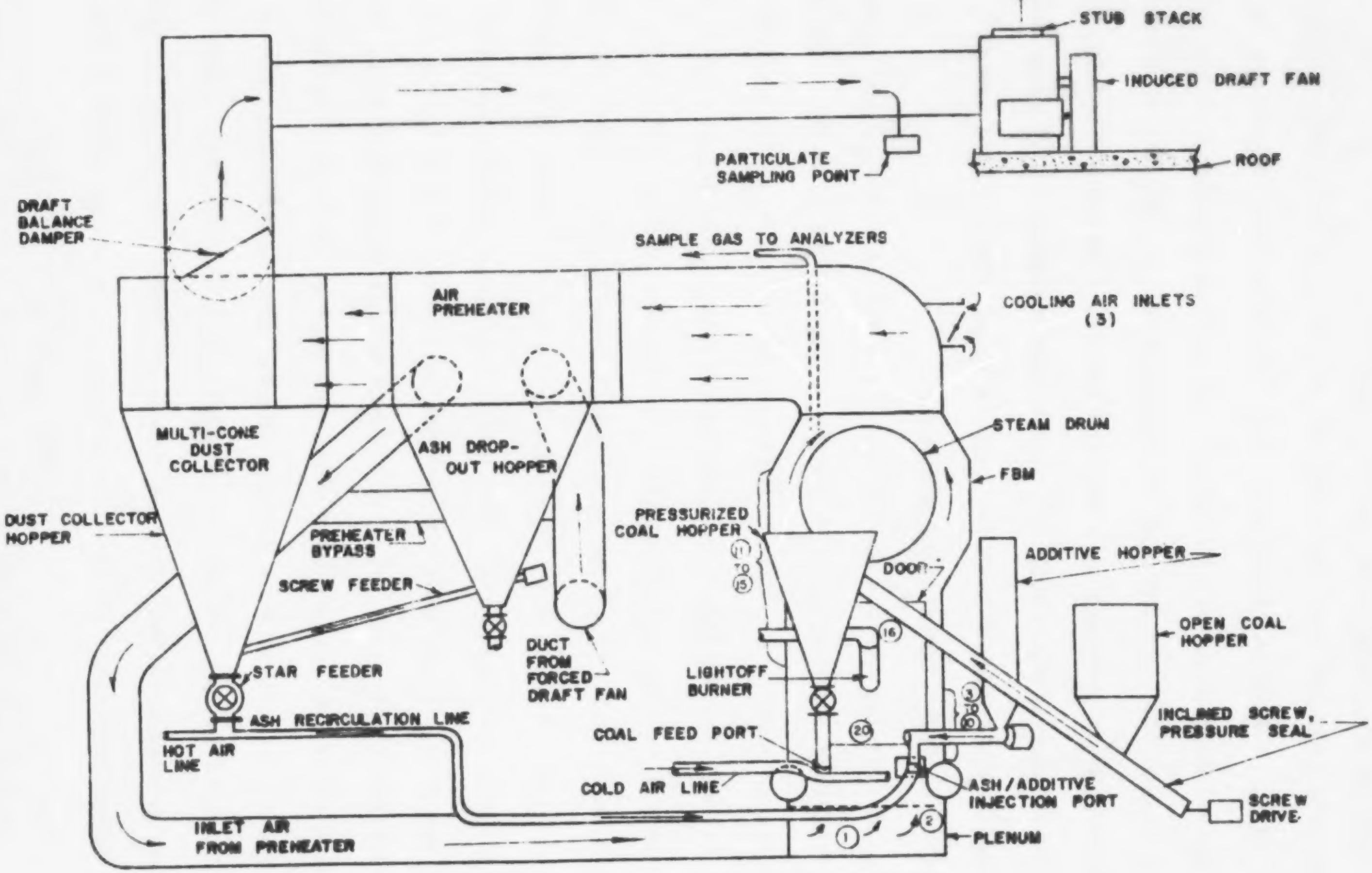
STUB STACK
INDUCED DRAFT FAN
ROOF
PARTICULATE SAMPLING POINT
DRAFT BALANCE DAMPER
SAMPLE GAS TO ANALYZERS
AIR PREHEATER
COOLING AIR INLETS (3)
STEAM DRUM
FBM
MULTI-CONE DUST COLLECTOR
ASH DROP-OUT HOPPER
DUST COLLECTOR HOPPER
PREHEATER BYPASS
PRESSURIZED COAL HOPPER
11 TO 15
ADDITIVE HOPPER
DOOR
SCREW FEEDER
16
OPEN COAL HOPPER
DUCT FROM FORCED DRAFT FAN
LIGHTOFF BURNER
STAR FEEDER
3 TO 10
ASH RECIRCULATION LINE
INCLINED SCREW, PRESSURE SEAL
20
COAL FEED PORT
HOT AIR LINE
COLD AIR LINE
ASH/ADDITIVE INJECTION PORT
SCREW DRIVE
INLET AIR FROM PREHEATER
1
2
PLENUM

Figure 71. Schematic diagram of PER-FBM test facility.

7.4.5 Babcock and Wilcox, Ltd. Renfrew Unit[44]

The data reported here was measured at the full-scale unit constructed by B&W, Ltd., in Renfrew, Scotland. This FBC units was constructed as a retrofit of an existing stoker-fired boiler. A schematic diagram of the unit is shown in Figure 72. The capacity is approximately 12 MW_t (40×10^6 Btu/hr).

Dried coal is conveyed to a storage bunker, from where it falls by gravity to a service hopper which supplies nine rotary feeders. Coal from these feeders is pneumatically conveyed into the bed via nine T-shaped feed points. Limestone and limited recycled fines added similarly.

The uncompartmented bed is 3.1 m × 3.1 m (10 ft × 10 ft) and operated with a fluidized depth of about 0.8 m to 0.9 m (2.6 ft to 3 ft). The distributor plate is made up of short stand pipes which admit air to the bed from the windbox below. The windbox is compartmented, thus allowing air to be shut off to sections of the bed independently, causing slumping and allowing turndown. There are three stand pipes for ash removal from the bed, although only one is generally used. The ash from this pipe falls into a cooler from where it is discharged via a rotary valve. Horizontal hairpin tubes are installed within the bed and provision is made for forced circulation of water from the boiler drum. For the first test series at a nominal 1.25 m/sec fluidizing velocity, two groups of boiler surface were provided, with an area of uncooled bed between. In total there were 10 tube loops. The boiler output was up to 10,500 kg/hr of steam. For the later tests at 2.5 m/sec, the number of tube loops was increased to 24. This increased the boiler output up to 21,000 kg/hr. About 50 percent of the heat absorption is accomplished in the submerged tubes.

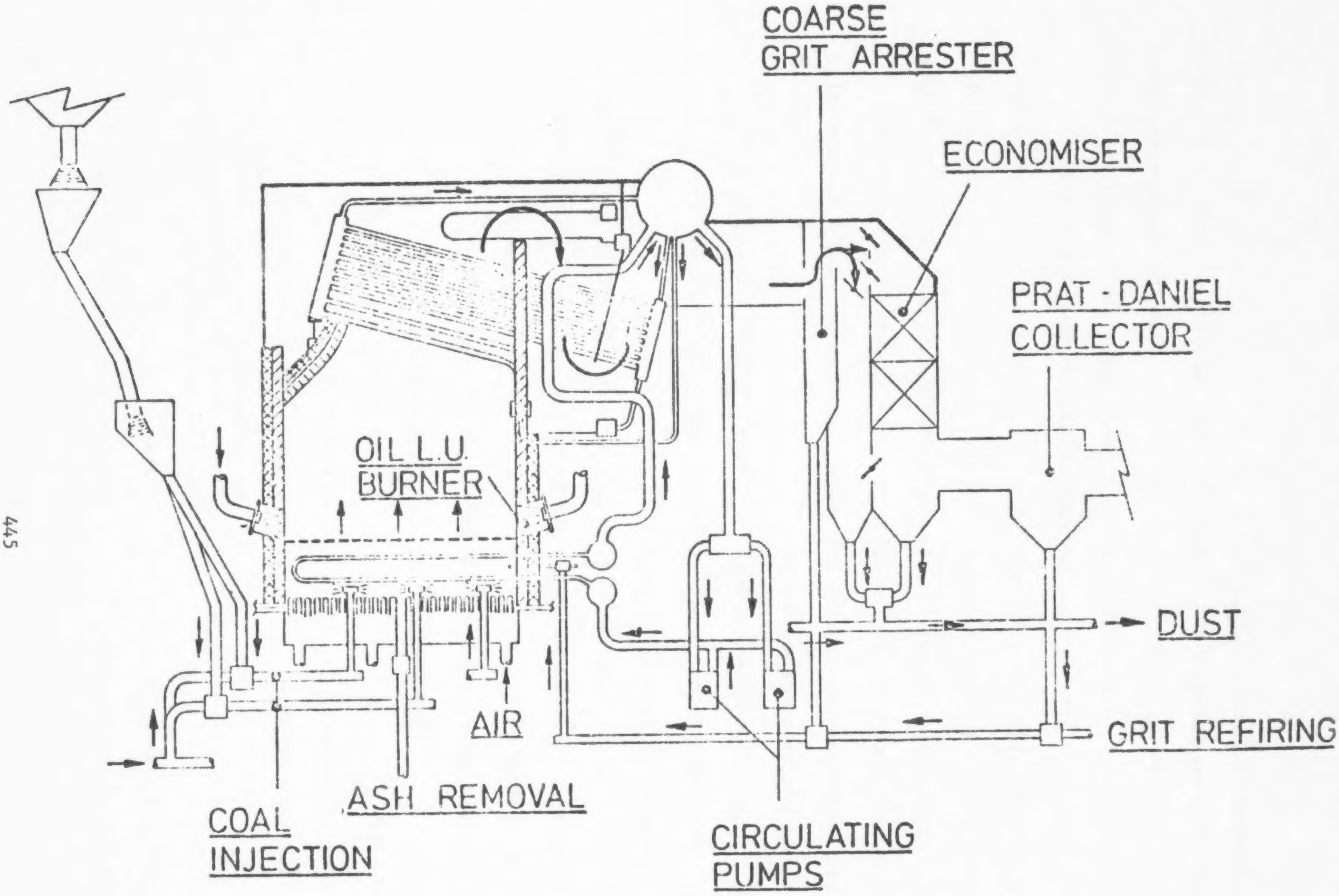
COARSE
GRIT ARRESTER
ECONOMISER
PRAT - DANIEL
COLLECTOR
OIL L.U.
BURNER
DUST
AIR
GRIT REFIRING
ASH REMOVAL
COAL
INJECTION
CIRCULATING
PUMPS

Figure 72. Schematic of the B&W, Ltd. designed Renfrew unit.

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7.4.6 FluiDyne 1.5 ft × 1.5 ft Unit[45]

This unit was designed and constructed after cold flow testing in a 0.6 m × 0.6 m (24 in. × 24 in.) plexiglass unit and is located at the Fluidyne Medicine Lake Test Facility. A schematic diagram of the pilot scale combustor is shown in Figure 73. Either inbed or abovebed feed is possible in this unit so that the effect of feed orientation on pollutant emissions can be observed. A primary cyclone is included and recycling is possible. Process air is raised from ambient temperature to 482°C (900°F) in a horizontal tube bundle heat exchanger located within the bed. It can be operated with or without preheated combustion air and uses a limestone or dolomite bed for SO_2 control. Other design operating parameters are:

- Superficial velocity, m/sec (ft/sec): 0.76 to 1.5 (2.5 to 5.0)
- Bed temperature, °C (°F): 788° to 898°C (1450° to 1650°F)

7.4.7 FluiDyne 3.3 ft × 5.3 ft Unit[46,47]

This unit was designed based on experience with the 1.5 ft × 1.5 ft unit. It is a vertical slice approximately one-third the size of a full-scale FBC module, as defined by FluiDyne. A schematic process diagram is shown in Figure 74. Design/operating conditions are listed below:

Test Combustor and Operating Conditions

Bed size	1.0 m × 1.62 m (40 in. × 64 in.)
Combustor pressure	Atmospheric
In-bed heat exchanger	Horizontal tube bundle for heating process air from ambient to 900°F (482°C) (full-scale tube length, diameter, packing density, and flow rate per tube).
Ignition burner fuel	Propane
Ignition burner location	Inlet to air distribution grid

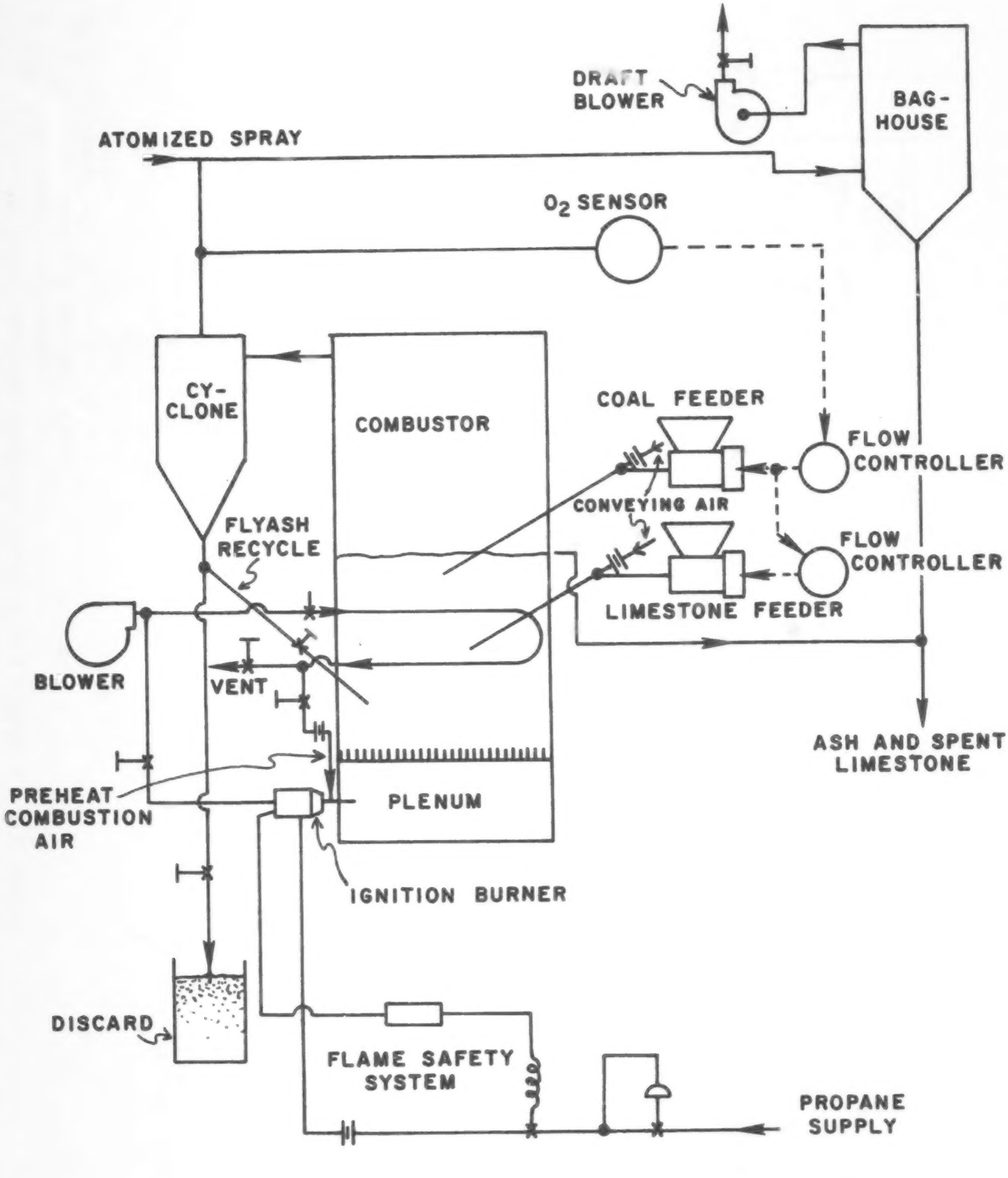

DRAFT BLOWER
BAG-HOUSE
ATOMIZED SPRAY
O2 SENSOR
CY-CLONE
COMBUSTOR
COAL FEEDER
FLOW CONTROLLER
CONVEYING AIR
FLOW CONTROLLER
FLYASH RECYCLE
LIMESTONE FEEDER
BLOWER
VENT
ASH AND SPENT LIMESTONE
PREHEAT COMBUSTION AIR
PLENUM
IGNITION BURNER
DISCARD
FLAME SAFETY SYSTEM
PROPANE SUPPLY

Figure 73. FluiDyne 1.5 ft × 1.5 ft pilot scale FBC combustor.

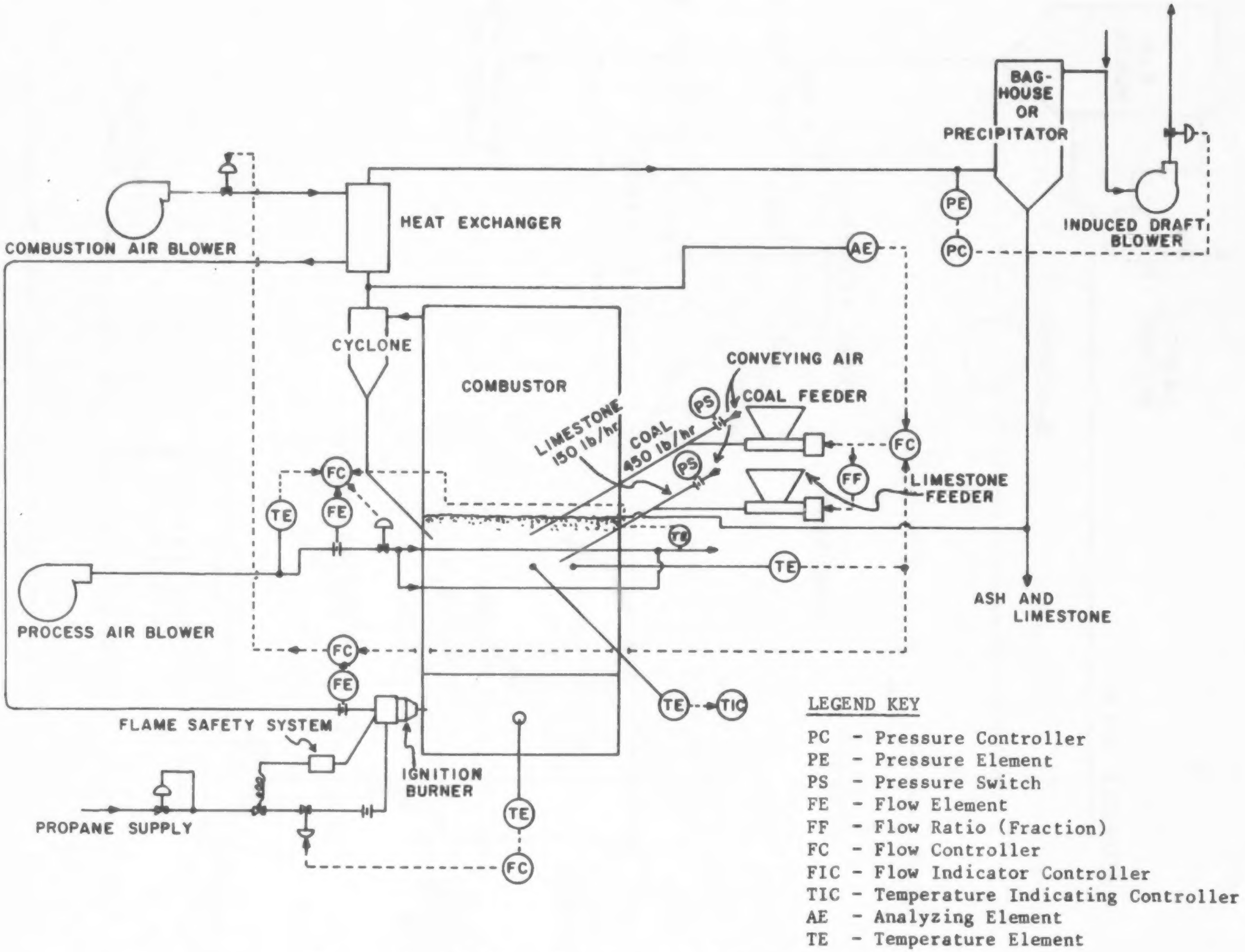

BAG-HOUSE OR PRECIPITATOR
COMBUSTION AIR BLOWER
HEAT EXCHANGER
INDUCED DRAFT BLOWER
PE
PC
AE
CYCLONE
COMBUSTOR
CONVEYING AIR
COAL FEEDER
PS
LIMESTONE 150 lb/hr
COAL 450 lb/hr
FC
FF
LIMESTONE FEEDER
FE
TE
PROCESS AIR BLOWER
ASH AND LIMESTONE
TIC
FLAME SAFETY SYSTEM
IGNITION BURNER
PROPANE SUPPLY
LEGEND KEY
PC - Pressure Controller
PE - Pressure Element
PS - Pressure Switch
FE - Flow Element
FF - Flow Ratio (Fraction)
FC - Flow Controller
FIC - Flow Indicator Controller
TIC - Temperature Indicating Controller
AE - Analyzing Element
TE - Temperature Element

Figure 74. FluiDyne 3.3 ft × 5.3 ft vertical slice FBC combustor.

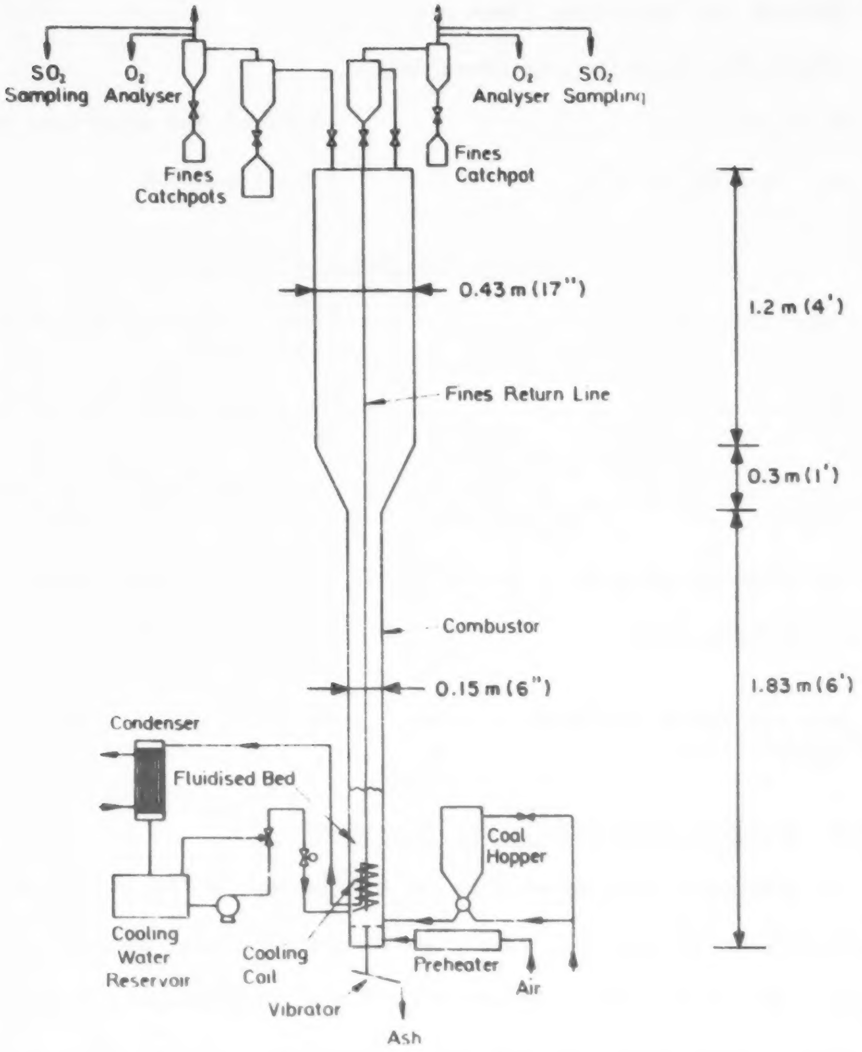
Superficial velocity	0.76 m/sec to 1.5 m/sec (2.5 ft/sec to 5.0 ft/sec)
Bed temperature	788° to 898°C (1450° to 1650°F)
Cyclone for recycling fines	
Limestone, dolomite, or inert bed	
Multipoint feed	(0.83 m^2 bed area/feed point)
Flue gas O_2 level	2 to 3 percent

System Flow Rates and Capacity

Combustion air	1180 to 2361 kg/hr (2600 to 5200 lb/hr)
Process air	0 to 5766 kg/hr (0 to 12,700 lb/hr)
Fuel feed rate	57 to 286 kg/hr (126 to 630 lb/hr)
Limestone feed rate	Varies with fuel sulfur
Total heat input	0.37 to 1.85 MW_t (1.25 to 6.3 × 10^6 Btu/hr)
Ash and spent limestone removal rate	Varies with fuel ash and sulfur

7.4.8 National Coal Board 6-in. Diameter Unit[48]

A schematic diagram of this unit is shown in Figure 75, with approximate dimensions. The unit was of circular cross-section, constructed of stainless steel. The whole combustor could be heated electrically by external wall heaters. These were used for startup and then to maintain a uniform temperature throughout the freeboard. Air was supplied from a plenum chamber, and passed through a distributor plate made from a drilled flat plate convered with three layers of 1 cm (3/8 in.) diameter alumina balls. The premixed coal/additive



SO_2 Sampling
O_2 Analyser
O_2 Analyser
SO_2 Sampling
Fines Catchpots
Fines Catchpot
0.43 m (17")
1.2 m (4')
Fines Return Line
0.3 m (1')
Combustor
0.15 m (6")
1.83 m (6')
Condenser
Fluidised Bed
Coal Hopper
Cooling Water Reservoir
Cooling Coil
Preheater
Air
Vibrator
Ash

Figure 75. National Coal Board 6-in. diameter FBC unit.

feed was pneumatically conveyed to the bed, which it entered tangentially, approximately 1.9 cm (3/4 in.) above the alumina balls. Excess heat was removed by a water-cooled metal coil immersed in the fluidized bed. The bed height was maintained constant by emptying surplus ash through a tube in the center of the distributor.

The gases leaving the combustor could be directed through two alternative cyclone systems, both comprising primary and secondary cyclones, for operation with or without fines recycle. With recycle, the primary cyclone was vertically above the bed and the fines were recycled via a dip-leg.

7.4.9 Argonne National Laboratories 6 in. Unit[33,34,35]

The Argonne 6-in. diameter atmospheric fluidized-bed combustor (shown in Figure 76) consisted of two vertical sections of stainless steel pipe. Four annular chambers (each 6.4 cm high) surround the lower section through which a mixture of water dispersed in air can be circulated to control heat removal in each zone. Figure 77 is a simplified piping diagram of the bench-scale equipment. Fluidizing air, after passing through a preheater at 538°C (1000°F) enters the reactor through a bubble-cap-type gas distributor mounted on the bottom flange of the reactor. Auxiliary heaters increase the inert-bed temperature to the coal ignition point. The coal, additive and recycled elutriated fines are entrained in transport air streams. Variable-drive volumetric screw feeders on scales are used to meter the solids into the transport air streams. The entrained solids are introduced into the fluidized-bed at a feed point just above the gas distributor. The off-gas from the reactor is passed through two high-efficiency cyclone separators in series and a cloth filter bag to effect separation of the solids from the gas stream. Provision was made for recycle of solids separated in the cyclone.

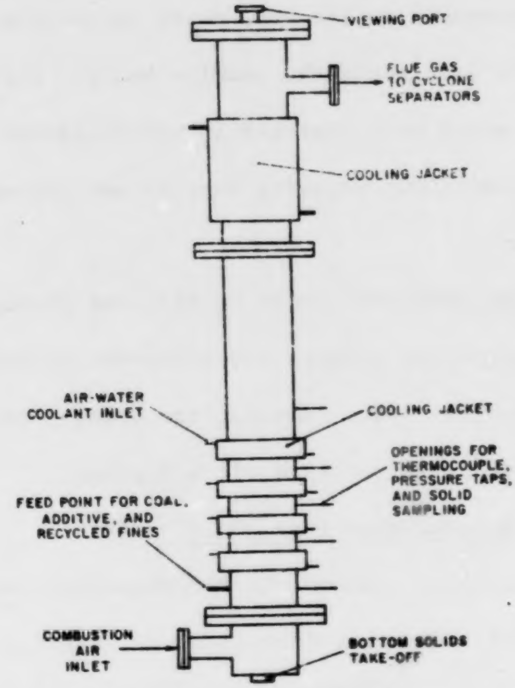

VIEWING PORT
FLUE GAS TO CYCLONE SEPARATORS
COOLING JACKET
AIR-WATER COOLANT INLET
COOLING JACKET
OPENINGS FOR THERMOCOUPLE, PRESSURE TAPS, AND SOLID SAMPLING
FEED POINT FOR COAL, ADDITIVE, AND RECYCLED FINES
COMBUSTION AIR INLET
BOTTOM SOLIDS TAKE-OFF

Figure 76. ANL 6-in. diameter bench-scale fluidized-bed combustion test unit.

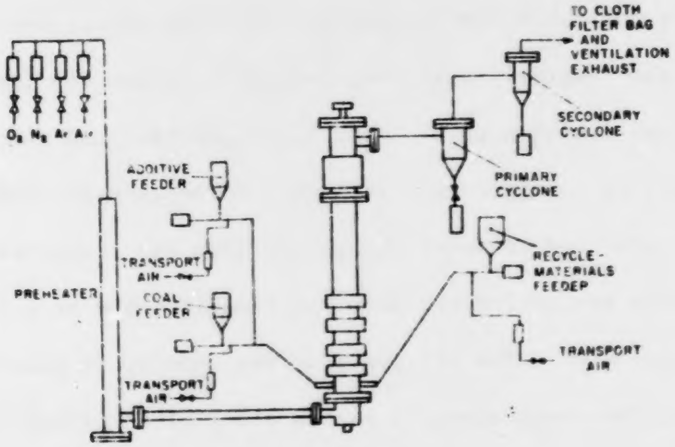

O_2
N_2
Ar
TO CLOTH FILTER BAG AND VENTILATION EXHAUST
SECONDARY CYCLONE
PRIMARY CYCLONE
ADDITIVE FEEDER
TRANSPORT AIR
PREHEATER
COAL FEEDER
TRANSPORT AIR
RECYCLE-MATERIALS FEEDER
TRANSPORT AIR

Figure 77. Overall diagram of ANL bench-scale equipment.

7.5 SUMMARY OF EMISSION SOURCE TEST DATA

The raw test data presented in Subsection 7.2 is summarized here in tabular form by pollutant emission; i.e., Table 92 presents SO_2 data, Table 93 presents NO_x data, and Table 94 presents particulate data. In most cases, emissions in terms of ng/J ($lb/10^6$ Btu) have been estimated by GCA from available data on flue gas concentrations and FBC operating conditions.

Test series have been grouped by coal type, sorbent type, or sorbent particle size. Throughout most test series, Ca/S molar feed ratios varied so that reporting average SO_2 emission reductions is meaningless. Therefore, only low and high SO_2 emissions recorded during each test series are reported, noting the applicable Ca/S ratios. This provides a more realistic basis for assessing those operating conditions which approached or supported the optional control levels being considered as part of this overall study. On this same basis, average emission values are not reported for NO_x or particulate emissions.

Table 95 shows the approximate average Ca/S ratios required to meet 75, 85, and 90 percent SO_2 reduction for the various sets of data in Section 7.0. These values were estimated by plotting the available data and interpolating for the optional SO_2 control levels. Extrapolation to 90 percent SO_2 control was necessary for the PER and the B&W 3 ft × 3 ft data. Variance within each set of data was usually dependent upon the type of limestone and the gas residence time used. In most cases, the units were operated at other than "best system" conditions. (See Subsection 7.6 for estimates of Ca/S requirements using "best system" conditions.) The points in the table represent an average trend in the data. Listed below are the maximum and minimum values extrapolated from the data.

TABLE 92. AFBC EMISSION SOURCE TEST DATA - SO_2

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions* mg/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average†					
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	28,907 (12,436)	3.46	7.29	Beckman Analyzing System (continuous)	2	-	0.47 - 0.48	133.3 (0.31)	133.3 (0.31)	†	94.3 - 94.4	NA	S	B&W Alliance, Ohio Ref. 1 Test 1-1	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 4.22
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	31,262 (13,440)	3.28	6.84	Beckman Analyzing System (continuous)	4	-	0.56 - 0.61	64.5 (0.15)	94.6 (0.22)	†	95.5 - 96.8	NA	S	B&W Alliance, Ohio Ref. 1 Test 1-2	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone 4.51 - 4.80
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	28,970 & 30,464 (12,464 & 13,106)	3.2 - 3.47	8.18 - 8.82	Beckman Analyzing System (continuous)	3	-	0.48 - 0.51	107.5 (0.25)	133.3 (0.31)	†	95.1 - 95.2	NA	S	B&W Alliance, Ohio Ref. 1 Test 1-3	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 4.06 - 4.59
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	31,589 (13,590)	3.29 - 3.39	5.93 - 6.83	Beckman Analyzing System (continuous)	4	-	0.38 - 0.41	98.9 (0.23)	116.1 (0.27)	†	94.0 - 94.9	NA	S	B&W Alliance, Ohio Ref. 1 Test 1-4	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 4.46 - 4.50
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	31,426 (13,520)	3.14	6.28	Beckman Analyzing System (continuous)	1	-	0.48	133.3 (0.31)	133.3 (0.31)	†	93.3	NA	S	B&W Alliance, Ohio Ref. 1 Test 1-5	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 4.2
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,506 (12,694)	3.48	6.68	Beckman Analyzing System (continuous)	1	-	0.46	541.8 (1.26)	541.8 (1.26)	†	78.8	NA	M	B&W Alliance, Ohio Ref. 2 Test 2-1	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.69
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,200 (12,600)	3.75 - 3.96	6.84 - 7.25	Beckman Analyzing System (continuous)	8	-	0.49 - 0.54	443 (1.03)	602 (1.40)	†	77.3 - 86.0	NA	L	B&W Alliance, Ohio Ref. 2 Test 2-2	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.4 - 3.2
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,784 (12,838)	3.21	6.32	Beckman Analyzing System (continuous)	2	-	0.56	1,152.4 (2.86)	1,152.4 (2.86)	†	55	NA	-	B&W Alliance, Ohio Ref. 2 Test 2-3	Ca/S = 0

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions[a] ng/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average[b]					
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,508 (12,686)	4.54	6.62	Beckman Analyzing System (continuous)	2	–	0.45 - 0.46	(1.79)	(1.95)	–	75.0 - 76.3	NA	M	B&W Alliance, Ohio Ref. 3 Test 3-2	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 1.87 - 1.98
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,894 (12,852)	3.69	6.05	Beckman Analyzing System (continuous)	3	–	0.41 - 0.48	(1.33)	(1.37)	–	76.17 - 76.79	NA	M	B&W Alliance, Ohio Ref. 3 Test 4-1 ABC	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.63 - 2.65
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,722 (12,778)	3.77	6.13	Beckman Analyzing System (continuous)	2	–	0.46 - 0.48	(1.33)	(1.34)	–	77.35 - 77.38	NA	M	B&W Alliance, Ohio Ref. 3 Test 4-1 DE	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.58 - 2.63
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,740 (12,786)	3.69	6.24	Beckman Analyzing System (continuous)	3	–	0.46	(1.26)	(1.45)	–	74.83 - 78.07	NA	M	B&W Alliance, Ohio Ref. 3 Test 4-1 FGH	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.63 - 2.66
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,663 (12,753)	3.87	6.32	Beckman Analyzing System (continuous)	3	–	0.46 - 0.48	589 (1.30)	585 (1.36)	–	77.60 - 78.54	NA	M	B&W Alliance, Ohio Ref. 3 Test 4-1 JJK	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S =
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,117 (12,518)	3.65	7.50	Beckman Analyzing System (continuous)	5	–	0.46 - 0.52	770 (1.79)	950 (2.21)	–	62.09 - 69.36	NA	–	B&W Alliance, Ohio Ref. 3 Test 4-1 LP	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.39 - 2.47
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,194 (12,551)	4.26	7.64	Beckman Analyzing System (continuous)	3	–	0.30	1,152 (2.68)	1,178 (2.74)	–	59.47 - 60.30	NA	–	B&W Alliance, Ohio Ref. 3 Test 4-2	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S =
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,324 (12,607)	4.22	6.57	Beckman Analyzing System (continuous)	3	–	0.44 - 0.50	598 (1.39	722 (1.68)	–	74.90 - 79.20	NA	–	B&W Alliance, Ohio Ref. 3 Test 4-3 AC	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.31 - 2.46

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions* mg/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average†					
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,168 (12,540)	6.14	7.14	Beckman Analyzing System (continuous)	2	–	0.48	615 (1.43)	662 (1.54)	–	76.67 - 78.32	NA	–	B&W Alliance, Ohio Ref. 3 Test 4-3 DE	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.56 - 2.57
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,368 (12,626)	3.89	7.51	Beckman Analyzing System (continuous)	5	–	0.46 - 0.48	357 (0.83)	477 (1.17)	–	81.0 - 86.56	NA	–	B&W Alliance, Ohio Ref. 3 Test 5-1	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 3.24 - 3.25
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	28,987 (12,462) 29,015 (12,474)	3.94 3.85	7.31 7.26	Beckman Analyzing System (continuous)	5	–	0.46 - 0.50	344 (0.80)	610 (1.42)	–	72.81 - 87.34	NA	–	B&W Alliance, Ohio Ref. 3 Test 5-2	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.47 - 3.61
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	28,791 (12,378)	4.12	7.68	Beckman Analyzing System (continuous)	3	–	0.40 - 0.41	808 (1.88)	1,040 (2.42)	–	63.61 - 71.78	NA	–	B&W Alliance, Ohio Ref. 3 Test 5-3	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.38 - 3.64
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	28,768 (12,368)	4.22	8.15	Beckman Analyzing System (continuous)	4	–	0.48 - 0.52	353 (0.82)	383 (0.89)	–	86.87 - 87.94	NA	I	B&W Alliance, Ohio Ref. 3 Test 6-1 AD	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 1.97 - 3.38
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,112 (12,516)	4.02	6.82	Beckman Analyzing System (continuous)	4	–	0.46	361 (0.84)	426 (0.99)	–	84.34 - 86.90	NA	I	B&W Alliance, Ohio Ref. 3 Test 6-1 EH	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 3.17 - 3.26
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW$_t$ (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,324 (12,607)	4.22	7.32	Beckman Analyzing System (continuous)	3	–	0.46 - 0.48	507 (1.18)	580 (1.35)	–	79.77 - 82.35	NA	M	B&W Alliance, Ohio Ref. 3 Test 6-1 IK	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 3.25 - 3.37

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics: Heat value kJ/kg (Btu/lb)	Fuel characteristics: % S	Fuel characteristics: % Ash	Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions[a] ng/J (lb/10^6 Btu): Low	Emissions[a]: High	Emissions[a]: Average[b]	Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,810 (12,816)	3.25	8.02	Beckman Analyzing System (continuous)	5	–	0.52	421 (0.98)	477 (1.11)	–	78.00 - 80.59	NA	M	B&W Alliance, Ohio Ref. 3 Test 6-1 LP	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.53 - 3.75
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,212 (12,559)	1.70	9.36	Beckman Analyzing System (continuous)	3	–	0.52 - 0.57	142 (0.33)	155 (0.36)	–	86.63 - 87.81	NA	I	B&W Alliance, Ohio Ref. 3 Test 6-2 ABC	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 4.67 - 4.74
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,170 (12,541)	2.53	8.82	Beckman Analyzing System (continuous)	5	–	0.52 - 0.55	236 (0.55)	267 (0.62)	–	84.73 - 86.24	NA	I	B&W Alliance, Ohio Ref. 3 Test 6-2 DM	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.36 - 3.67
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,815 (12,818)	2.27	8.14	Beckman Analyzing System (continuous)	4	–	0.52 - 0.55	254 (0.59)	288 (0.67)	–	81.07 - 83.22	NA	M	B&W Alliance, Ohio Ref. 3 Test 6-2 TL	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.44 - 2.73
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,536 (12,698)	2.58	9.35	Beckman Analyzing System (continuous)	5	–	0.48 - 0.52	370 (0.86)	408 (0.95)	–	76.70 - 78.79	NA	M	B&W Alliance, Ohio Ref. 3 Test 6-2 MQ	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.59 - 2.69
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,743 (12,787)	2.87	8.30	Beckman Analyzing System (continuous)	6	–	0.41 - 0.46	417 (0.97)	563 (1.31)	–	70.73 - 78.35	NA	M	B&W Alliance, Ohio Ref. 3 Test 6-3 AF	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.27 - 2.40
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (25 × 10^6 Btu) as tested	AFBC Limestone Addition	29,770 (12,779)	2.18	8.04	Beckman Analyzing System (continuous)	6	–	0.43 - 0.44	224 (0.52)	245 (0.57)	–	83.16 - 84.75	NA	M	B&W Alliance, Ohio Ref. 3 Test 6-3 GJ	Sorbent = <9,525 μm (3/8 in. × 0) Lowellville Limestone Ca/S = 2.96 - 3.03

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions* ng/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average†					
0.9 m × 0.9 m (3 ft × 3 ft) 1.75 MW (6×10^6 Btu/hr) as tested	AFBC Limestone Addition	29,375 (12,586)	3.06	9.32	Dupont Model 411 UV light absorption	6	10	0.17 - 0.20	400 (0.93)	696 (1.62)	†	66 - 81	NA	M	B&W Alliance, Ohio Ref. 6	Sorbent = 6,350 μm (1/4 in. × 0) Lowellville Limestone Ca/S = 0.58 - 2.71
		29,375 (12,586)	2.86	9.43		3	7	0.17 - 0.42	1,251 (2.91)	1,685 (3.92)	†	13 - 36		-		Sorbent = 6,350 μm (1/4 in. × 0) Lowellville Limestone Ca/S = 1.57 - 1.81
		29,484 (12,676)	2.86	9.43		6	10	0.16 - 0.17	585 (1.36)	1,582 (3.68)	†	18 - 70		-		Sorbent = 2,380 μm (8 mesh) × 0 Lowellville Limestone Ca/S = 1.11 - 3.51
		29,486 (12,676)	2.86	9.43		4	8.5	0.16 - 0.17	666 (1.56)	1,582 (3.68)	†	18 - 66		-		Sorbent = 1,000 μm (16 mesh) × 0 Lowellville Limestone Ca/S = 0.87 - 3.49
		29,486 (12,676)	2.86	9.43		3	8.5	0.13 - 0.19	791 (1.84)	1,040 (2.42)	†	46 - 59		-		Sorbent = pulverized Lowellville Limestone Ca/S = 2.05 - 2.38
		29,484 (12,676)	2.86	9.43		3	12.5	0.16 - 0.21	353 (0.82)	507 (1.18)	†	74 - 82		M		Sorbent = 44 μm (325 mesh) × 0 Hydrated Lime ($Ca(OH)_2$) Ca/S = 1.68 - 2.18
		29,115 (12,517)	3.12	9.76		1	5.5	0.17	825 (1.92)		†	61		-		Sorbent = pulverized Lowellville Limestone Ca/S = 3.76
		29,115 (12,517)	3.12	9.76		1	1.5	0.18	1,543 (3.59)		†	28		-		Sorbent = 4 μm (325 mesh) Hydrated Lime Ca/S = 0.99
		29,115 (12,517)	3.12	9.76		3	7	0.16 - 0.18	391 (0.91)	1,148 (2.67)	†	46 - 82		M		Sorbent = Greer limestone 3 sizes (8M, 16M, Pulv) Ca/S = 2.70 - 3.94
		29,115 (12,517)	3.12	9.74		3	9	0.13 - 0.17	460 (1.07)	765 (1.78)	†	66 - 79		M		Sorbent Grove Limestone 3 sizes (8M, 16M, Pulv) Ca/S = 3.93 - 4.93

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions[a] ng/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average[e]					
0.91 × 0.46 (3 × 1.5)	AFBC Limestone Addition	35,062 (15,074)	2.8	13.5	IR	9	4	0.53 - 0.77	30 (0.07)	796 (1.85)	e	50 - 98	NA	Stringent	NCB-CRE Test 1 Ref. 5 & 6	Limestone 18 210 μm median Ca/S = 1.2 - 3.3
	AFBC Limestone Addition	33,437 (14,375)	1.3	18.2	IR	5	½	0.26 - 0.58	218 (.51)	482 (1.12)	e	38 - 72	NA	-	NCB-CRE Test 3 Ref. 5 & 6	U.K. Limestone 300 - 400 μm median Ca/S = 1.8 - 3.0
	AFBC Limestone Addition	35,062 (15,074)	2.8	13.5	IR	8	½	0.26 - 0.76	9 (0.02)	1,054 (2.45)	e	34 - 100	NA	Stringent	NCB-CRE Test 2 & 5 Ref. 5 & 6	Limestone 18 350 - 500 μm median Ca/S = 1.0 - 6.0
	AFBC Dolomite Addition	35,062 (15,074)	2.8	13.5	IR	6	1	0.53 - 1.76	16 (0.06)	447 (1.06)	e	72 - 99	NA	Stringent	NCB-CRE Test 4 Ref. 5 & 6	Dolomite 1337 100 - 130 μm median Ca/S = 1.6 - 3.1
	AFBC Dolomite Addition	35,062 (15,074)	2.8	13.5	IR	5	½	0.36 - 0.88	112 (0.26)	575 (1.33)	e	64 - 93	NA	Stringent	NCB-CRE Test 6 Ref. 5 & 6	Dolomite 1337 875 - 1025 μm median Ca/S = 2.5 - 5.4
1.5 ft × 6 ft	AFBC Dolomite Addition	Ohio No. 8 unwashed 30,084 (12,936)	4.3	10.7	IR	8	1	0.13 - 0.26	1,369 (3.2)	2,931 (6.8)	e	2.6 - 56.4	NA	-	PER FBM Alexandria, Virginia Ref. 7	Dolomite 1337 raw -2,830 + 1,410 μm Ca/S ratio: 1.15 - 1.75
	AFBC Limestone Addition	Ohio No. 8 unwashed 30,084 (12,934)	4.3	10.7	IR	3	1	0.13 - 0.26	1,634 (3.8)	2,150 (5.0)	e	28.2 - 45	NA	-	PER FBM Alexandria, Virginia Ref. 7	Natural mine limestone - 2,830 + 1,410 μm Ca/S ratio: 1.11 - 1.70
	AFBC Dolomite Addition	Ohio No. 8 unwashed 30,084 (21,936)	4.3	10.7	IR	5	1	0.13 - 0.26	1,376 (3.2)	2,150 (5.0)	e	28.2 - 56	NA	-	PER FBM Alexandria, Virginia Ref. 7	Dolomite 1359 Hydrate -44 μm Ca/S ratio: 0.72 - 0.98
		Ohio No. 8 unwashed 30,084 (12,936)	4.3	10.7	IR	4	1	0.13 - 0.26	776 (1.8)	1,077 (2.5)	e	64.2 - 74.2	NA	-	PER FBM Alexandria, Virginia Ref. 7	Dolomite 1337 raw -44 μm Ca/S ratio: 1.7 for low and high values. One test using Ca/S = 1.9 reported midrange SO$_2$ reduction of 70.9 percent.

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics: Heat value kJ/kg (Btu/lb)	% S	% Ash	Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions^e ng/J (lb/10^6 Btu): Low	High	Average^f	Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
1.5 ft × 6 ft	AFBC Limestone Addition	Ohio No. 8 unwashed 30,084 (12,934)	6.5	10.7	IR	6		0.13 - 0.26	782 (1.6)	1,206 (2.8)	e	60 - 76		-	PER FBM Alexandria, Virginia Ref. 7	Limestone 1359 raw -44 µm Ca/S ratio: 1.7 - 2.0
	AFBC Limestone Addition	Ohio No. 8 washed 31,820 (12,934)	2.6	7.2	IR	6		0.13 - 0.26	523 (1.3)	1,226 (2.9)	-	25 - 68		-	PER FBM Alexandria, Virginia Ref. 7	Limestone 1337 Hydrate -44 µm Ca/S ratio: 1.17 - 1.46
1.5 ft × 6 ft	AFBC Dolomite Addition	Ohio No. 8 washed 31,820 (13,680)	2.6	7.2	IR	4		0.13 - 0.26	267 (0.63)	567 (1.3)	e	69.3 - 81.9	NA	Moderate	PER FBM Alexandria, Virginia Ref. 7	Dolomite 1337 raw -44 µm Ca/S ratio: 2.2 - 2.4 (high and low emissions were reported at Ca/S = 2.2)
	AFBC Dolomite Addition	Ohio No. 8 washed 31,820 (13,680)	2.6	7.2	IR	5		0.13 - 0.26	464 (1.1)	623 (1.44)	f	61.9 - 71.6	NA	-	PER FBM Alexandria, Virginia Ref. 7	Limestone 1359 raw -44 µm Ca/S ratio: 1.6 - 2.6
	AFBC Dolomite Addition	Ohio No. 8 washed 31,820 (13,680)	2.6	7.2	IR	4		0.13 - 0.26	619 (1.5)	917 (1.9)	f	50 - 61.6	NA	-	PER FBM Alexandria, Virginia Ref. 7	Limestone 1359 Hydrate -44 µm Ca/S ratio: 1.4 - 1.6
1.5 ft × 6 ft	AFBC Limestone Addition	Sewickley coal Sewickley coal	4.1-4.5	-	IR	7		0.13 - 0.26	679 (1.58)	967 (2.25)	f	-	NA		PER FBM Alexandria, Virginia Ref. 8	Germany Valley Limestone Ca/S ratio: 2.8 - 4.4 1500 - 2500 µm median
	AFBC Limestone Addition		4.1-4.5	-	IR	6		0.13 - 0.26	473 (1.1)	1,071 (2.49)	f	-	NA			Greer Limestone Ca/S ratio: 2.9 - 3.5 1500 - 2500 µm median

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics: Heat value kJ/kg (Btu/lb)	Fuel characteristics: % S	Fuel characteristics: % Ash	Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions* ng/J (lb/10⁶ Btu): Low	Emissions* ng/J (lb/10⁶ Btu): High	Emissions* ng/J (lb/10⁶ Btu): Average†	Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
10 ft × 10 ft	AFBC Limestone Addition	–	5.3	–	–	3	–	–	–	–	–	65 - 95	NA	S	B&W, Ltd. Renfrew, Scotland Ref. 9	Limestone A Ca/S = 0.8 - 3.2
10 ft × 10 ft	AFBC Limestone Addition	–	5.5	–	–	4	–	–	–	–	–	30 - 95	NA	S	B&W, Ltd. Renfrew, Scotland Ref. 9	Limestone B Ca/S = 1.8 - 6.0
1.5 ft × 1.5 ft	AFBC Limestone Addition	–	4.8	–	Beckman Model 865	3	–	0.67	–	–	–	60 - 76	NA	M	FluiDyne 18" × 18" at FluiDyne Medicine Lake Test Facility Ref. 10	Above-bed feed No recycle Illinois Limestone Ca/S Ratio = 3
1.5 ft × 1.5 ft	AFBC Limestone Addition	–	4.8	–	Beckman Model 865	3	–	0.67	–	–	–	84 - 90	NA	S	FluiDyne 18" × 18" at FluiDyne Medicine Lake Test Facility Ref. 10	In-Bed feed No recycle Illinois Limestone Ca/S Ratio = 3
1.5 ft × 1.5 ft	AFBC Limestone Addition	–	4.8	–	Beckman Model 865	4	–	0.67	–	–	–	80 - 85	NA	S	FluiDyne 18" × 18" at FluiDyne Medicine Lake Test Facility Ref. 10 & 12	Above-bed feed With recycle Illinois Limestone Ca/S Ratio = 3
1.5 ft × 1.5 ft	AFBC Limestone Addition	–	4.8	–	Beckman Model 865	2	–	0.67	–	–	–	93 - 95	NA	S	FluiDyne 18" × 18" at FluiDyne Medicine Lake Test Facility Ref. 10 & 12	In-bed feed With recycle Illinois Limestone Ca/S Ratio = 3
3.3 ft × 5.3 ft	AFBC Dolomite Addition	–	3.6	–	Beckman Model 865 Dupont Model 411	1	–	0.86	–	–	–	87.2	NA	I	FluiDyne 18" × 18" at FluiDyne Medicine Lake Test Facility Ref. 10 & 12	Run No. 35 Above-bed feed With recycle
3.3 ft × 5.3 ft	AFBC Dolomite Addition	–	3.6	–	Beckman Model 865 Dupont Model 411	1	500	0.85 - 2.0	–	–	–	80	NA	M	FluiDyne 40" × 64" Verticle Slice Combustor Ref. 11 & 12	500 Hour test In-bed feed With recycle Ca/S = 1.7 - 2.4

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions* ng/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average†					
6 in. Diameter	AFBC Limestone Addition	Illinois Coal	4.4	11.8	Iodine method Haffmann-Braum I.R. and H_2O_2	4	-	0.67	-	-	-	0-94	NA	S	NCB 6 in. Diameter FBC Ref. 12	U.K. Limestone Ca/S = 0-3.2
6 in. Diameter	AFBC Limestone Addition	Welbeck Coal	1.3	18.2	Iodine method Haffmann-Braum I.R. and H_2O_2	9	-	0.67 - 1.00	-	-	-	0-97	NA	S	NCB 6 in. Diameter FBC Ref. 12	U.K. Limestone Ca/S = 0-2.9
6 in. Diameter	AFBC Limestone Addition	Park Hill	2.5	16.5	Iodine method Haffmann-Braum I.R. and H_2O_2	7	-	0.67	-	-	-	0-84	NA	S	NCB 6 in. Diameter FBC Ref. 12	U.K. Limestone Ca/S = 0-2.6
6 in. Diameter	AFBC Limestone Addition	Pittsburgh	2.8	13.5	Iodine method Haffmann-Braum I.R. and H_2O_2	6	-	0.67	-	-	-	0-88	NA	S	NCB 6 in. Diameter FBC Ref. 12	U.K. Limestone Ca/S = 0-3.1
6 in. Diameter	AFBC Limestone Addition	Pittsburgh	2.8	13.5	Iodine method Haffmann-Braum I.R. and H_2O_2	6	-	0.67	-	-	-	0-91	NA	S	NCB 6 in. Diameter FBC Ref. 12	Limestone 18 Ca/S = 0-2.6
6 in. Diameter	AFBC Limestone Addition	Welbeck	1.3	18.2	Iodine method Hoffmann-Braum I.R. and H_2O_2	1	-	0.67	-	-	-	80	NA	M	NCB 6 in. Diameter FBC Ref. 12	Limestone 18 Ca/S = 1.9
6 in. Diameter	AFBC Limestone Addition	Illinois Coal	4.4	11.8	Iodine method Haffmann-Braum I.R. and H_2O_2	10	-	0.67	-	-	-	0-93	NA	S	NCB 6 in. Diameter FBC Ref. 12	Limestone 1359 Ca/S = 0-3.6

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont duration (hrs)	Gas residence time (sec)	Emissions* ng/J (lb/10^6 Btu)			Range of control § (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average†					
6 in. Diameter	AFBC Limestone Addition	28,126 (12,092)	4.63	12.39	Beckman I.R. Model 315A	17	-	0.33	430 (1.00)	3,295 (7.66)	-	0-87	NA	I	ANL 6 in. Unit CC Test Series Ref. 13-16	Limestone 1337 100 - 1200 μm Ca/S = 0-5.1
6 in. Diameter	AFBC Limestone Addition	28,482 (12,245)	4.84	13.13	Beckman I.R. Model 315A	18	-	0.22 and 0.67	375 (0.87)	2,785 (6.98)	-	18 - 89	NA	I	ANL 6 in. Unit SACC Test Series Ref. 13-16	Limestone 1359 25m 600, 1200, 1400 μm Ca/S = 1.0-3.0
6 in. Diameter	AFBC Limestone Addition	28,482 (12,245)	4.84	13.13	Beckman I.R. Model 315A	20	-	0.22 and 0.67	170 (0.40)	3,400 (7.91)	-	0 - 95	NA	S	ANL 6 in. Unit SA Test Series Ref. 13-16	Limestone 1359 25 and 103 μm Ca/S = 0-4.2
6 in. Diameter	AFBC Limestone Addition	28,482 (12,245)	4.84	13.13	Beckman I.R. Model 315A	19	-	0.67-0.71	140 (0.32)	2,820 (6.56)	-	17 - 96	NA	S	ANL 6 in. Unit BC Test Series Ref. 13-16	Tymochtee Dolomite Limestone 1359 Limestone 1360 Limestone 1337 25 - 615 μm Ca/S = 0-2.6
6 in. Diameter	AFBC Limestone Addition	28,475 (12,242)	4.84	13.13	Beckman I.R. Model 315A	6	-	0.71	305 (0.71)	1,905 (4.43)	-	44 - 91	NA	S	ANL 6 in. Unit AR-1 Test Series Ref. 13-16	Limeston 1339 490 μm Ca/S = 2.5
6 in. Diameter	AFBC Limestone Addition	28,290 (12,163)	3.7 and 4.1	10.8 and 12.08	Beckman I.R. Model 315A	12	-	0.41-0.77	615 (1.43)	1,815 (4.22)	-	38 - 79	NA	M	ANL 6 in. Unit AMER Test Series Ref. 13-16	Limestone 1359 555 - 609 μm Ca/S = 1.05-2.99
6 in. Diameter	AFBC Limestone Addition	27,463 (11,807)	1.28	18.07	Beckman I.R. Model 315A	3	-	0.77-0.80	170 (0.39)	430 (1.00)	-	55 - 82	NA	M	ANL 6 in. Unit BRIT Test Series Ref. 13-16	B-Sonk Limestone 440 μm Ca/S = 1.2-3.65

(continued)

TABLE 92 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Gas residence time (sec)	Emissions[*] ng/J (lb/10 Btu)			Range of control (%)	Design control efficiency of device	Maximum optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash					Low	High	Average[†]					
6 in. Diameter	AFBC Limestone Addition	28,290 (12,163)	4.14	12.08	Beckman IR Model 315 A	2	–	0.77	320 (0.74)	940 (2.18)	–	66 - 68	NA	–	ANL 6 in. Unit AM-BRIT Series Ref. 13-16	Limestone 1359 and B-Sonk 550 μm & 440 μm Ca/S = 1.05 - 2.9
6 in. Diameter	AFBC Limestone Addition	Humphrey Coal	2.4	–	Beckman IR Model 315 A	13	–	0.71 - 0.83	–	–	–	39 - 47	NA	–	ANL 6 in. Unit Bump Series Ref. 13-16	Limestone 1359 Ca/S = 0.94 - 4.58
6 in. Diameter	AFBC Limestone Addition	–	2.4	–	Beckman IR Model 315 A	9	–	0.26 - 0.83	–	–	–	0	NA	–	ANL 6 in. Unit HP Series Ref. 13-16	No Sorbent Addition Ca/S = 0
6 in. Diameter	AFBC Limestone Addition	28,290 (12,163)	3.7	10.85	Beckman IR Model 315 A	10	–	0.66 - 0.83	–	2,615 (6.08)	–	–	NA	–	ANL 6 in. Unit Peabody Series Ref. 13-16	Limestone 1359 Ca/S = 0 - 4.54

[*] Variation for each test group correlates with the Ca/S ratio used.

[†] An average is inappropriate for these tests due to the variation in test conditions for each test series.

[‡] Test duration varied from 2 to 4 hours for each set of conditions after steady state conditions were reached.

[§] Test duration determined by time required to reach steady state conditions.

Note: NA = Not Applicable.

[#] High value measured using -125 μm limestone at Ca/S ratio of 3.6.

TABLE 93. AFBC EMISSION SOURCE TEST DATA - NO_x

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J ($lb/10^6$ Btu)			Range of control (%)	Design control efficiency of device	Range of optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low	High	Average^a					
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW_t (24×10^6 Btu/hr)	AFBC	29,194 (12,551)	4.24	7.64	Beckman I.R.	3	-	112 (0.26)	120 (0.28)	*	NA	NA	S	B&W Alliance, Ohio Ref. 4 Test 4-2	Fuel N = 1.13%
		29,324 (12,607)	4.22	6.57		5	-	77 (0.18)	112 (0.26)	*	NA	NA	S	B&W Alliance, Ohio Ref. 4 Test 4-3	Fuel N = 1.22%
		29,168 (12,540)	4.14	7.14											
		29,368 (12,626)	3.89	7.51		5	-	150 (0.35)	163 (0.38)	*	NA	NA	S	B&W Alliance, Ohio Ref. 4 Test 5-1	Fuel N = 1.14%
		28,987 (12,462)	3.94	7.31		2	-	185 (0.43)		*	NA	NA	S	B&W Alliance, Ohio Ref. 4 Test 5-2 A-B	Fuel N = 1.22%
		29,015 (12,474)	3.85	7.24		4	-	150 (0.35)		*	NA	NA	S	B&W Alliance, Ohio Ref. 4 Test 5-2 C-F	Fuel N = 1.03%
		28,768 (12,368)	4.22	8.15		16	-	95 (0.22)	116 (0.27)	*	NA	NA	S	B&W Alliance, Ohio Ref. 4 Test 6-1	Fuel N = 1.22%, 1.24% 1.31%
		29,810 (12,816)	3.25	8.02											
		29,112 (12,516)	4.02	6.82											

(continued)

TABLE 93 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Range of optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low	High	Average*					
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW, (24 × 10^6 Btu/hr)	AFBC	29,212 (12,559)	1.70	9.36	Beckman I.R.	17	-	90 (0.21)	155 (0.36)	*	NA	NA	S	B&W Alliance Ohio Ref. 4 Test 6-2	Fuel N = 1.31%, 1.32%, 1.34%
		29,170 (12,541)	2.53	8.82											
		29,815 (12,818)	2.27	8.14											
		29,536 (12,698)	2.58	9.35											
		29,743 (12,787)	2.87	8.50		10	-	138 (0.32)	146 (0.34)	*	NA	NA	S	B&W Alliance Ohio Ref. 4 Test 6-3	Fuel N = 1.23% 1.24%
		29,770 (12,799)	2.18	8.04											
0.9 m × 0.9 m (3 ft × 3 ft) 1.75 MW (6 × 10^6 Btu/hr) as tested		29,375 (12,629)	3.04	9.32	Teco Model 10A Chemiluminescence	4	10	73 (0.17)	133 (0.31)	*	NA	NA	S	B&W Alliance, Ohio Ref. 5	Fuel N = 0.86%
		29,375 (12,629)	2.86	9.43		3	7	47 (0.11)	291 (0.51)	*			I-S		Fuel N = 0.86%
		29,484 (12,676)	2.86	9.43		6	10	155 (0.36)	228 (0.53)	*			I-S		Fuel N = 0.76%
		29,484 (12,676)	2.86	9.43		4	8.5	0 (0)	236 (0.52)	*			I-S		Fuel N = 0.76%
		29,484 (12,676)	2.86	9.43		3	8.5	125 (0.29)	185 (0.43)	*			S		Fuel N = 0.76%
		29,484 (12,676)	2.86	9.43		3	12.5	129 (0.30)	219 (0.51)	*			I-S		Fuel N = 0.76%
		29,115 (12,517)	3.12	9.74		1	5.5	150 (0.35)					S		Fuel N = 1.23%
		29,115 (12,517)	3.12	9.74		1	1.5	189 (0.44)					S		Fuel N = 1.23%

(continued)

TABLE 93 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10^6 Btu)			Range of control (%)	Design control efficiency of device	Range of optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low	High	Average*					
0.9 m × 0.9 m (3 ft × 3 ft) 1.75 MW (6 × 10^6 Btu/hr) as tested	AFBC	29,115 (12,517)	3.12	9.74	Teco Model 10A Chemiluminescence	3	7	176 (0.41)	262 (0.61)	*	NA	NA	M-S	B&W Alliance, Ohio Ref. 5	Fuel N = 1.23%
		29,115 (12,517)	3.12	9.74		3	9	150 (0.35)	219 (0.51)	*			I-S		Fuel N = 1.23%
0.91 m × 0.46 m (36 in. × 18 in.)	AFBC	Pittsburgh coal 35,062 (15,073)	2.8	13.5	BCURA NO_x box	6	†	126 (0.29)	225 (0.52)	*	NA	NA	I-S	NCB-CRE 36 in. × 18 in. Ref. 4	Coal size < 1,680 µm Dolomite 1337 < 1,680 µm
		Pittsburgh coal 35,062	2.8	13.5	BCURA NO_x box	5	†	191 (0.44)	226 (0.53)	*			I-S	NCB-CRE 36 in. × 18 in.	Coal size < 3,175 µm Dolomite 1337 < 3,175 µm
		Pittsburgh coal 35,062	2.8	13.5	BCURA NO_x box	8	†	191 (0.44)	323 (0.75)	*			None-S	NCB-CRE 36 in. × 18 in.	Coal size < 3,175 µm Limestone 18 < 3,175 µm
1.5 ft × 6 ft	AFBC	Ohio No. 8 unwashed 30,084 (12,934)	4.5	10.7	IR	11		87 (0.2)	216 (0.50)	*	NA	NA	S	PER FBM Alexandria, Virginia Ref. 8,9	Coarse (-2,830 + 1,410 µm) limestone addition
		Ohio No. 8 unwashed 30,084 (12,934)	4.5	10.7		11		91 (0.21)	187 (0.44)	*			S		Fine (-44 µm) limestone addition
		Ohio No. 8 washed 31,820 (13,680)	2.6	7.2		19		107 (0.25)	228 (0.53)	*			I-S		Fine (-44 µm) limestone addition
		Washed and unwashed Ohio No. 8				25		107 (0.25)	228 (0.53)	*			I-S		All tests without sorbent addition

(continued)

TABLE 93 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10⁶ Btu)			Range of control (%)	Design control efficiency of device	Range of optional control level supported	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low	High	Average*					
1.5 ft × 6 ft	AFBC	Washed and unwashed Ohio No. 8			IR	41		87 (0.20)	216 (0.50)	*			S	PER FBM Alexandria, Virginia Ref. 8,9	All tests with sorbent addition
10 ft × 10 ft	AFBC	-	5.5	-	‡	11		65 (0.15)	200 (0.45)	*	NA	NA	S	Renfrew, Scotland Ref. 10	Estimated from ppm reported
3.3 ft × 5.3 ft	AFBC	Illinois No. 6 -	3.6		Beckman IR Model 865	1	500	159 (0.37)	236 (0.55)	*	NA	NA	I-S	Fluidyne Ref. 11, 12	Express air from 30% to 130%
6 in. diameter	AFBC	28,482 (12,245)	4.84	13.13	Beckman IR	57	-	115 (0.29)	460 (1.07)	*	NA	NA	None-S	ANL Ref.14-17 SA Series, SACC Series and BC Series	Fuel N = 1.11%
		28,290 (12,163)	3.7	10.85	Beckman IR	33	-	5 (0.07)	585 (0.90)	*	NA	NA	None-S	ANL Ref.14-17 AR AMER Peabody Series	Fuel N = 1.18%
		Humphrey -	2.4	-	Beckman IR	22	-	195 (0.45)	390 (0.91)	*	NA	NA	None-S	ANL Ref.14-17 HUMP & HP Series	

*Averages are inappropriate for these tests due to the variation in test conditions within each series.

†Test duration varied from 2 to 4 hours for each set of test conditions.

‡Chemiluminescence.

Note: NA = Not applicable.

TABLE 94. AFBC EMISSION SOURCE TEST DATA - PARTICULATE LOADING TO FINAL CONTROL DEVICE

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10^6 Btu)			Range of control necessary to meet S control† (%)	Range of control necessary to meet I control‡ (%)	Range of control necessary to meet M control‡ (%)	Opacity (%)	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low*	High*	Average						
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW (25 × 10^6 Btu/hr) as tested	Primary cyclone	29,506 (12,694)	3.48	6.68	As described in Section 7.3.1	1	-	2,750 (6.4)		†	99.5	98.4	96.1	NR	B&W Alliance, Ohio Ref.1,2,3 Test 2-1	These emission rates represent loadings to a final particulate control device
		29,500 (12,600)	3.75 2.96	6.84 7.25		8	-	2,710 (6.3)	4,260 (9.9)	†	99.5-99.7	98.4-99.0	96.1-97.5			
		29,784 (12,838)	3.21	6.32		2	-	1,850 (4.3)		†	99.3	97.7	94.2			
		29,508 (12,686)	4.54	6.62		2	-	3,224 (7.5)		†	99.6	98.7	96.7			Cyclone outlet loadings greater than 2150 ng/J (5.0 lb/10^6 Btu) usually occurred when primary collection efficiency was reported below 75 percent
		29,500 (12,680)	3.76	6.75		11	-	3,323 (7.73)-	6,453 (15.01)	†	99.6-99.8	98.7-99.3	96.8-98.3	NR		
		29,194 (15,551)	4.24	7.64		3		3,130 (7.28)-	3,147 (7.32)	†	99.6	98.6	96.6			
		29,200 (12,560)	4.15	6.86		5		3,203 (7.45)-	3,431 (7.98)	†	99.6	98.7	96.6-96.9			
		29,368 (12,626)	3.94	7.51		5		2,042 (4.75)-	2,068 (4.81)	†	99.4	97.9	94.7-94.8			
		28,987 (12,462)	3.94	7.25		6		770 (1.79)-	1,367 (3.18)	†	98.3-99.1	94.4-96.9	86.0-92.1			
		28,791 (12,378)	4.12	7.68		3		2,129 (4.98)-	2,206 (5.13)	†	99.4	98.0-98.1	94.9-95.1			

(continued)

TABLE 94 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10^6 Btu)			Range of control necessary to meet S control‡ (%)	Range of control necessary to meet I control‡	Range of control necessary to meet M control‡	Opacity (%)	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low*	High*	Average						
1.9 m × 1.9 m (6 ft × 6 ft) 7 MW (25×10^6 Btu/hr) as tested	Primary cyclone	29,324 (12,607)	3.74	7.50	As described in Section 7.3.1	16	-	1,638 (3.81)	2,184 (5.08)	†	99.2-99.4	97.4-98.0	93.4-95.1	NR		
		29,536 (12,698)	2.27	8.82		12	-	1,961 (4.56)-	3,276 (7.62)	†	99.3-99.6	97.8-98.7	94.5-96.7			
		29,743 (12,787)	2.87	8.50		10	-	3,603 (8.38)-	3,770 (8.77)	†	99.6-99.7	98.8-98.9	97.0-97.1			
0.9 m × 0.9 m (3 ft × 3 ft) 1.75 MW (6×10^6 Btu/hr) as tested	Integral low efficiency collector	29,375 (12,629)	3.04	9.32	As described in Section 7.3.2	4	10	2,683 (6.24)	3,156 (7.34)	†	99.5-99.6	98.4-98.6	96.0-96.6	NR	B&W Alliance, Ohio Ref. 4	These emission rates represent loadings to a final particulate control device. Loadings are high because limited freeboard and low primary removal efficiency permitted substantial carryover.
		29,375 (12,629)	2.86	9.43		3	7	2,253 (5.24)	3,375 (7.85)	†	99.4-99.6	98.1-98.7	95.2-96.8			
		29,484 (12,676)	2.86	9.43		6	10	2,878 (6.74)	3,689 (8.58)	†	99.6-99.7	98.5-98.8	96.3-97.0			
		29,484 (12,676)	2.86	9.43		4	8.5	3,078 (7.16)	4,170 (9.70)	†	99.6-99.8	98.6-99.0	96.5-97.4			
		29,484 (12,676)	2.86	9.43		3	8.5	5,434 (12.64)	5,765 (13.41)	†	99.8	99.2-99.3	98.0-98.1			
		29,484 (12,676)	2.86	9.43		3	12.5	6,453 (15.01)	7,145 (16.62)	†	99.8	99.3-99.4	98.3-98.5			
		29,115 (12,517)	3.12	9.74		1	5.5	7,825 (18.20)		†	99.8	99.5	98.6			
		29,115 (12,517)	3.12	9.74		1	1.5	4,970 (11.56)		†	99.7	99.1	97.8			
		29,115 (12,517)	3.12	9.74		3	7	3,637 (8.46)	10,623 (24.71)	†	99.6-99.9	98.8-99.6	97.0-99.0			
		29,115 (12,517	3.12	9.74		3	9	3,457 (8.04)	15,215 (35.39)	†	99.6-99.9	98.8-99.7	96.9-99.3			

(continued)

TABLE 94 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10^6 Btu)			Range of control necessary to meet S control‡ (%)	Range of control necessary to meet I control‡ (%)	Range of control necessary to meet M control‡ (%)	Opacity (%)	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low*	High*	Average						
18 in. × 72 in.	Integral primary multicone collector	Ohio No. 8 unwashed 30,084 (12,934)	4.5	10.7	Isokinetic sampling at one point downstream of primary multicone collector	2	-	318* (0.74)	696* (1.62)	†	95.9-98	86.5-93.8	66.2-84.6	NR	PER FBM Alexandria, Virginia Ref. 7,8	Dolomite 1337 raw -44 μm % reduction not reported since final fly ash control is required after primary removal
		Ohio No. 8 unwashed 30,084 (12,934)	4.5	10.7		2	-	374* (0.87)	679* (1.58)	†	96.6-98.1	86.5-93.7	71.3-84.2	NR	PER FBM Alexandria, Virginia Ref. 7,8	Limestone 1359 raw -44 μm
		Ohio No. 8 washed 31,820 (13,680)	2.6	7.2		2	-	396* (0.92)	4.94* (1.15)	†	96.7-97.4	89.1-91.3	72.8-78.3	NR	PER FBM Alexandria, Virginia Ref. 7,8	Limestone 1337 hydrate -44 μm
		Ohio No. 8 washed 31,820 (13,680)	2.6	7.2		2	-	383* (0.89)	602* (1.4)	†	96.6-97.9	88.8-92.9	71.9-82.1	NR	PER FBM Alexandria, Virginia Ref. 7,8	Limestone 1337 raw -44 μm
		Ohio No. 8 washed 31,820 (13,680)	2.6	7.2		2	-	374* (0.87)	598* (1.39)	†	96.6-97.8	88.5-92.8	71.3-82.0	NR	PER FBM Alexandria, Virginia Ref. 7,8	Limestone 1359 raw -44 μm

(continued)

TABLE 94 (continued)

Actual boiler size	Control method	Fuel characteristics			Test method	Number of tests	Longest cont. duration (hrs)	Emissions ng/J (lb/10⁶ Btu)			Range of control necessary to meet S control‡ (%)	Range of control necessary to meet I control‡ (%)	Range of control necessary to meet M control‡ (%)	Opacity (%)	Reference unit I.D. and location	Remarks
		Heat value kJ/kg (Btu/lb)	% S	% Ash				Low*	High*	Average*						
18 in. × 72 in.		Ohio No. 8 washed 31,820 (13,680)	2.6	7.2		2		327* (0.76)	473* (1.1)		96.1-97.3	86.8-90.9	67.1-77.3	NR	PER FBM Alexandria, Virginia Ref. 7,8	Limestone 1359 hydrate -44 μm

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[*] High value measured during fine sorbent addition; low value measured with no sorbent addition.

[†] An average is inappropriate for these tests due to the variation in test conditions for each test series.

[‡] S = 0.03 lb/10⁶ Btu (12.9 ng/J)
I = 0.10 lb/10⁶ Btu (43 ng/J)
M = 0.25 lb/10⁶ Btu (107.5 ng/J)

Note: NA = Not applicable; NR = Not reported.

TABLE 95. AVERAGE Ca/S REQUIREMENTS TO MEET THREE LEVELS OF CONTROL. EXTRAPOLATED FROM TABLES 81 THROUGH 91[*]

No.	Unit ID	range of gas residence time (sec)	Range of sorbent particle size, µm and limestone type	Ca/S required to meet 75 percent reduction	85	90
Table	B&W 6 ft × 6 ft	0.30 - 0.61	< 9,525 Lowellville Limestone	2.1	3.3	3.8
Table	B&W 3 ft × 3 ft	0.13 - 0.24	6,350 × 0 to pulverized Lowellville, $Ca(OH)_2$, Greer and Grove limestones	3.5	4.0	4.3
Table	NCB-CRE 36 in. × 18 in.	0.26 - 1.76	100 - 1,000 Limestone 18, U.K. limestone and dolomite 1337	2.6	3.1	3.3
Table	Per FBM[*]	0.13 - 0.26	2,830 - 44 raw and hydrated dolomite 1337; raw and hydrated limestone 1359	2.1	-	-
Figure	B&W Ltd Renfrew	-	Limestone B	1.6	2.0	2.3
			Limestone A	4.2	4.9	5.3
Tables	NCB 6 in. diameter	-	U.K. limestone Limestone 1359 Limestone 18	2.2	2.7	2.9
Table	ANL 6 in. diameter	0.22 - 0.83	100 - 1,200 Dolomite 1337, Limestone 1360, Limestone 1359, Tymochtee Dolomite, B-Sonk	2.8	3.3	3.6

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[*] FluiDyne results are not reported. A single Ca/S ratio of 3.0 was used in the 1.5 ft × 1.5 ft data reviewed, and all levels of SO_2 control were supported depending on operating conditions. In the FluiDyne vertical slice testing reviewed, 80 percent SO_2 removal was obtained at a Ca/S ratio of 2.4 and 1.7, the lower value corresponding to a longer gas residence time. In one other 3.3 ft × 5.3 ft test run (No. 35), 87 percent SO_2 reduction was achieved at a Ca/S ratio of 2.4.

[†] Insufficient data to extrapolate to emission levels of 85 and 90 percent reduction.

Percentage SO_2 removal	Minimum Ca/S	Maximum Ca/S
75	1.5	5.1
85	1.6	5.2
90	2.2	5.6

In general, the objective of experimental programs to date has been to characterize emissions as a function of FBC operating conditions. The research has been primarily exploratory in nature; the FBC units were small, and much of the testing preceeded the proposal of the EPA Reference Sampling Methods. As a result, the EPA Reference Methods have not been employed extensively on FBC units in the past. In addition, previous testing has not generally been conducted at FBC operating conditions designed exclusively for the most cost-effective means of environmental control. In the very near future, most FBC testing programs will define, in more detail, performance of FBC at more optimal conditions for pollution control and will include the use of the EPA Reference Methods.

The emissions data summarized in the tables are discussed below. No attempt is made to compare the results of one experimental program to another: i.e., PER versus B&W, because test conditions and unit designs vary widely. Therefore, the discussion is limited to the results determined by each investigator, and methods by which the efficiency of pollution abatement could have been enhanced.

7.5.1 Babcock and Wilcox Company 6 ft × 6 ft Unit

B&W ran a series of tests during 1978 and 1979 to demonstrate the SO_2 control capability of fluidized-bed combustion in their 6 ft × 6 ft unit. The test series show that 75, 85, and 90 percent SO_2 reduction is achievable. Greater than 90 percent SO_2 removal was also achieved using Ca/S ratios greater than 4. The results were impressive considering the apparently large limestone particle size and relatively short gas phase residence time which averaged about 0.5 sec.

The NO_x data reported in Table 93 all meet 215 ng/J (0.5 lb/10^6 Btu) the optional stringent emission guideline. In fact, of 56 data points all but 1 are under 172 ng/J (0.4 lb/10^6 Btu) and two-thirds are under 129 ng/J (0.3 lb/10^6 Btu). The data is promising as the B&W unit is one of the larger units for which data is available and, thus, best represents industrial boiler capacity. The gas residence times are also slightly lower than those recommended for best systems, thus, there is a potential for decreasing the NO_x emissions even further.

Other variables during testing were: temperature, from 834°C (1533°F) to 899°C (1650°F); gas residence time from 0.30 sec to 0.61 sec (as compared to ~0.7 sec, which is currently thought to be appropriate for effective SO_2 control); fuel ranging in heating value from 28,768 kJ/kg (12,368 Btu/lb) to 31,589 kJ/kg (13,590 Btu/lb); sulfur content from 1.70 percent to 4.54 percent; and ash content from 5.93 percent to 9.36 percent.

The B&W test series also included two tests during which there was no sorbent addition. Comparing the dust loading at the cyclone inlet of these two tests, 4,686 ng/J (10.9 lb/10^6 Btu), to those tests which did have sorbent addition, between 6,535 ng/J (15.2 lb/10^6 Btu) and 11,092 ng/J (25.8 lb/10^6 Btu) shows the relative amount of particulate elutriation which can be attributed to sorbent addition. If fines recycling were greater, this impact could perhaps be lessened.

Complete recycling was not possible during this testing because the current system is designed to recycle only about one-fourth of the carryover. If more efficient recycling were possible, higher SO_2 removals could be anticipated at the Ca/S ratios used, due to higher utilization of calcium.

7.5.2 Babcock and Wilcox 3 ft × 3 ft Unit

The 31 tests which B&W reported have been summarized into 10 categories, although no two tests in any one category were actually run under exactly the same conditions. The type of fuel, type of limestone, and particle size were used to distinguish the categories. Within these categories the variation in SO_2 emissions is dependent primarily on the Ca/S molar feed ratio and the limestone particle size. For example, using Lowellville limestone at 6,350 µm × 0 (1/4 in. × 0), B&W found that SO_2 reduction increased from 66 to 81 percent as Ca/S molar feed ratio was increased from 0.58 to 2.71. (Increasing the Ca/S ratio further would produce even greater sulfur capture according to the trend in the test data.) The data also show that as the particle size is decreased from 2,380 µm (8 mesh) to 1,000 µm (16 mesh) (for the Grove and Greer limestones), the sulfur retention increases slightly. Further decrease in particle size should also increase sulfur retention. However, B&W data shows that the actual sulfur removal rate is lower for the case of the pulverized limestones. This decrease is accounted for by particle elutriation which was reported to be extremely high compared to that noted during addition of larger particles. The small particles probably elutriated from the bed before reaction with SO_2 could occur. If the captured carryover had been recycled, better SO_2 capture may have occurred. By reducing gas velocity, thus reducing particle elutriation and increasing gas residence time, a marked improvement in sulfur retention may have been possible.

The main objective of the testing was to assess the effect of operating variations on boiler performance and emissions reductions. B&W was also attempting to reduce required boiler size by increasing gas velocity to 8 ft/sec with a bed height limit of about 1-1/2 ft. Although the reported SO_2

emissions from this test series generally meet only a moderate SO_2 control level at best with the Ca/S ratios used by B&W, this is not surprising since the gas residence time was fairly low (generally 0.2 sec or less) and the sorbent particle size was either fairly large (1000 to 6350 μm) or else was so small (~44 μm) that the sorbent elutriated from the bed before it could completely react. As mentioned in several previous sections, current theory suggests that improved performance could be achieved if the gas residence time were increased by a factor of 3 or 4 (e.g., 0.6 to 0.8 sec) and limestone particles on the order of 500 μm in the bed were used.

Another major contributor to the lower SO_2 removal efficiencies in the 3 ft × 3 ft unit is the unit's low freeboard (allowing carryover of sorbent particles before they had adequate time to react with SO_2), combined with the lack of recycle of the carryover back to the bed (so that once elutriated, the sorbent particles did not have any further opportunity to react).

The tradeoffs between designing larger boilers with lower fluidizing velocities but enhanced SO_2 capture and staying with current FBC designs are currently being studied.*

All of the NO_x data reported supports an optional intermediate standard (258 ng/J, 0.6 lb/10^6 Btu). Seventy percent of the results support the optional stringent level of control (215 ng/J, 0.5 × 10 Btu). There is no apparent experimental variable which had a predominant influence on NO_x emission levels in this test program.

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*Current designs in fluidized-bed combustion tend to stress crushed stone and high fluidizing velocity. The impact on overall FBC design features of switching to pulverized stone and lower fluidizing velocity is currently being studied by GCA, Gilbert/Commonwealth, and Westinghouse under EPA Contract.

The particulate data reported represents the loading at the inlet of a wet scrubber; hence it is uncontrolled. The 3 ft × 3 ft unit at Babcock and Wilcox did not include a primary cyclone which is a part of most FBC systems and used to recycle bed carryover. In addition, the freeboard of this unit is very limited, resulting in carryover of relatively large particles which are "splashed" out of the bed, and would not normally be entrained. The low freeboard, together with the high gas velocities used, contribute to the high dust loadings prior to the scrubber. Therefore, it is impractical to use the data to project the control efficiency required of an additional final particulate control device. However, this data is discussed in Section 2.0 because available particulate data for FBC systems is limited.

7.5.3 National Coal Board 3 ft × 1.5 ft Test Unit

As shown in Table 83 and 84, the test series with Pittsburgh coal and limestone 18 at a median particle size of 210 μm consisted of nine runs with Ca/S ratios ranging from 1.2 to 3.3. Sulfur control ranged from 50 percent with a Ca/S ratio of 2.2 up to 98 percent with a Ca/S ratio of 3.3. The lowest control level of 50 percent appears out of line because four other runs at a Ca/S ratio of 2.2 showed retention of sulfur at 76, 81, 83, and 84 percent. In looking more closely at the data, two factors could have contributed to the low removal in the one test; a low bed temperature of 749°C; and a large percentage of very fine particles in the bed material.[26] The relatively small sorbent particle size (about 210 μm mass mean) used in this entire test series, combined with the general absence of recycle, could have resulted in elutriation of much of the sorbent before it had a chance to fully react (at the velocities of 0.9 to 1.2 m/sec (3 to 4 ft/sec) being used) with no opportunity (through recycle) for additional reaction time in the bed. This test series was operated, however, with gas residence times (0.53 to 0.77 sec) in the range being suggested in this report for effective SO_2 removal.

For the series with Welbeck coal and U.K. limestone at 300 to 400 μm median particle feed size, sulfur control was not high at the Ca/S ratios used. Retention ranged from 38 percent up to 72 percent with Ca/S ratio at 1.8 and 3.0, respectively. The trend of the data indicates that if a sufficiently high Ca/S ratio had been used, support of optional SO_2 emissions levels may have been achieved. The gas phase residence time for this test series was quite low, around 0.3 sec. A residence time of 0.6 sec or greater would have increased SO_2 retention. In this test series, recycle of the primary cyclone catch would probably have improved SO_2 capture.

A series of eight runs with Pittsburgh coal and limestone 18 at 350 to 450 μm average particle size resulted in sulfur retention ranging from 34 up to 99+ percent. Ca/S ratios ranged from 1.1 to 6.0. Two additional tests were run at similar conditions with a significantly reduced limestone size (<125 μm). These tests indicate control of 49 percent at a Ca/S ratio of 1 with the 125 μm sorbent as compared with 34 percent control at similar conditions with the larger particle size. Variations in the series also show that the best capture efficiency was achieved with the highest gas residence times. For example, the one run with a residence time of >0.5 sec gave 65 percent SO_2 removal at a Ca/S ratio of 1.7. The other data points are at conditions reflecting lower residence times (0.26 to 0.45 sec) and thus do not portray the calcium utilization that might be possible with residence times of 0.6 to 0.7 sec. Again, recycle of the cyclone catch may be expected to have improved the SO_2 capture, especially in the 125 μm cases.

Pittsburgh coal and dolomite 1337 at median feed size of 100 to 125 μm achieved SO_2 control levels from 72 to 99 percent. In most of these tests, gas residence time was about 0.5 sec and recycle was not employed. In this series Ca/S ratios of 2.6 to 3.4 supported control levels of 75 to 85 percent. In the final run, the Ca/S ratio was set at 1.6 and fines were recycled to the bed (the only test conducted with recycle). In addition, in the last test, a gas residence time of 1.76 sec was used, which is much greater than in the other tests and which is possibly greater than would be cost-effective in commercial practice. An SO_2 emission reduction of 99 percent was achieved at the very attractive Ca/S ratio of 1.6, showing the acute impact of long gas residence time, and the use of fine sorbent with recycle, on the amount of sorbent needed to meet high SO_2 capture constraints. The "best system" of SO_2 control in AFBC considered in this report envisions a shorter gas residence time (~0.67 sec) than considered in this last test (and hence a smaller boiler). Sorbent particle size envisioned is also more coarse (500 μm surface mean), resulting in less grinding cost for commercial applications. Recycle of primary cyclone catch is also envisioned in the "best system." A commercial AFBC system that employed fine sorbent (125 μm) and high recycle rates, as suggested by this last test, could be attractive in commercial practice, but has not been considered explicitly in this report due to the limited data available on this method of operation.

Another series was run with Pittsburgh coal and dolomite 1337 at a median size of 875 to 1000 μm. In one run at a Ca/S ratio of 2.6, sulfur retention was 64 percent. For the rest of the series, retention was in the intermediate to stringent control range (87 to 93 percent) with Ca/S ratios between 5.0 and 5.4. However, only one measurement showed greater than 90 percent SO_2 reduction.

Once again, the gas residence time is an obvious factor in the calcium utilization. Using approximately the same Ca/S (5.3 and 5.2) at gas residence times of 0.49 and 0.88 sec, sulfur capture was 88 and 93 percent, respectively.

In reviewing the NCB SO_2 removal data, it appears that the relatively low percentage of SO_2 removal in many cases is a combined effect resulting from the low residence times, small limestone particle sizes combined with relatively high velocities, and the absence of primary bed recycle. For example, the median limestone sizes generally range from 200 to 500 µm at an 8 ft/sec gas velocity and 100 µm at 4 ft/sec; the sorbent is probably being blown out of the bed before it can react completely. Without recycle, there is no chance for further reaction.

The data that exist are not inconsistent with achieving SO_2 removal at levels between 85 to 90 percent. Extrapolating the trends in the data indicate that under suitable operating conditions, these removal levels could be achieved (see Figure 88).

As shown in Tables 83 and 84, emissions of NO_x were reported for three test series: (1) Pittsburgh coal with limestone 18 at <3,175 µm; (2) Pittsburgh coal with dolomite 1337 at <3,175 µm; and (3) Pittsburgh coal with dolomite 1337 at <1,680 µm. The low and high emissions in ng/J for these tests are, respectively, 191 to 323; 191 to 226; and 126 to 225. None of the parameters investigated have a strong influence on NO_x emissions.

7.5.4 Pope, Evans, and Robbins

The SO_2 emission test data measured at the FBM is grouped in Table 92 by coal type, limestone type, and sorbent particle size. Addition of coarse sorbents provided a maximum SO_2 reduction of 54.5 percent at a Ca/S ratio of 1.75 when unwashed high sulfur coal was burned. SO_2 reduction was increased to 74 percent at a Ca/S ratio of 1.7 when raw dolomite 1337 was fed at -44 μm. The same reduction was attained using raw limestone 1359 at -44 μm and a Ca/S ratio of 2.0.

Burning washed medium sulfur coal indicates similar SO_2 reductions as a function of limestone type, Ca/S ratio, and limestone particle size, although coarse sorbents were not tested with washed coal. The maximum SO_2 reduction measured was 82 percent at a Ca/S ratio of 2.2 using raw dolomite 1337. This was the only case in which an optional SO_2 control level was supported. Gas residence times were generally very low (about 0.15 sec) and could account for results which do not appear optimum. It must be stressed, however, that when these tests were conducted, support of specific SO_2 control levels was not the objective. By extrapolating the data exhibited in the table, one can speculate that increased Ca/S ratios and increased gas residence time would support intermediate and stringent SO_2 control levels. The summary table also shows that the hydrated sorbents did not exhibit greater SO_2 removal capability than the raw forms.

In interpreting the PER data, it is critical to note that: (1) gas residence times were normally quite low, typically 0.15 to 0.25 sec; (2) sorbent particle size was either very coarse (-2800, +1400 μm) limiting available reaction surface area, or so fine (-44 μm) that it elutriated very rapidly at the high gas velocities being employed (3 m/sec (10 ft/sec) or higher); (3) the freeboard above the fluidized bed was very limited, allowing significant carryover; and (4) in general, the carryover captured by the cyclone was not recycled, except in a few cases. All of the factors together contributed to the relatively low SO_2 removals observed in the FBM.

Referring back to Table 84, it is possible to assess SO_2 test results based on continuous IR analysis as compared to wet chemical analysis according to EPA Reference Method 6. In all cases, SO_2 emissions in terms of ppm are very close for the two techniques. Differences in reported emissions are within the range expected based on the precision of either of the two analysis techniques.

Later testing results burning Sewickley coal using Greer and Germany Valley limestone (see Tables 85 and 92) showed fairly high SO_2 emissions in terms of ng/J ($lb/10^6$ Btu) although fairly high Ca/S ratios were used. Gas residence times of about 0.2 sec were used which are not as high as would be desirable for effective SO_2 removal. It is not possible to calculate reliable values of percentage reduction due to lack of data, but the maximum reduction using Greer limestone at a Ca/S ratio of 3.5 is probably in the range of 80 to 85 percent. PER has noted that Germany Valley limestone has a higher calcium content, but Greer limestone has a more favorable internal structure and more favorable overall kinetics.

The average NO_x emission measured during all the Pope, Evans, and Robbins FBM testing reported in 1970 was approximately 275 ppm or 175 ng/J (0.4 $lb/10^6$ Btu). NO_x data was not included in the presentation of results during combustion of Sewickley coal in the FBM. Table 93 shows low and high NO_x values recorded during combustion of unwashed and washed Ohio coals with coarse and fine sorbent addition. The range of NO_x measured is also shown for the overall testing with and without sorbent feed.

Comparison of NO_x measurements based on IR analysis and methods similar to EPA Reference Method 7 (see Table 84 and emissions reported in ppm) illustrates good agreement between the two techniques. Only three of the 16 comparisons differ by as much as a factor of 2. Most values are within a range of

±10 percent. The larger differences were noted in the first test runs, and then good agreement was demonstrated as experimentation continued.

Table 94 shows particulate emissions downstream of the multiclone collector based on the washed and unwashed coal and the different sorbents. Each test series includes a dust loading measurement with sorbent feed and without sorbent feed. In all cases, the higher emissions level was associated with addition of finely divided sorbent. With sorbent addition, the data suggests that final fly ash control of greater than 90 percent efficiency is required to achieve an intermediate optional control level of 43 ng/J (0.1 $lb/10^6$ Btu).

7.5.5 Babcock and Wilcox, Ltd.

Because limited data were available, a summary tabulation of emissions data from the Renfrew unit is not included. However, some useful information can be extracted from the graphical results presented earlier in Subsection 7.2.2.

Figure 57 illustrates SO_2 reduction as a function of Ca/S molar feed ratio, using two different limestones. It is important to note that SO_2 emissions reductions greater than 90 percent were achieved burning high sulfur (5.5 percent) coal using a Ca/S ratio of about 2.5 with a more reactive sorbent, but a Ca/S ratio of about 5 would be necessary if the less reactive sorbent were used.* The curves also illustrate that laboratory scale tests accurately predict SO_2 reduction in a full-scale industrial boiler.

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*No details were provided on the specific differences between the two types of sorbent.

In Figure 58, NO_x emissions during combustion of a coal containing 1.1 percent nitrogen are shown as a function of bed temperature. The analyses were done by the chemiluminescence method. The maximum emission level of 325 ppm (corrected to stoichiometric conditions) is equivalent to approximately 195 ng/J (0.45 lb/10^6 Btu), which supports the optional stringent NO_x control level under consideration.

7.5.6 FluiDyne 1.5 ft × 1.5 ft Unit

FluiDyne reported the results of SO_2 emission testing in this unit at the Fifth International Conference on Fluidized-Bed Combustion. The data is important because it demonstrates the effect of feed orientation and primary recycle. Without primary recycle, SO_2 removal efficiency with abovebed feed is inferior to removal efficiency attained with inbed feed at the same Ca/S ratio (approximately 3). This is caused by the lower sorbent/SO_2 reaction time available due to rapid elutriation of small sorbent particles without subsequent reinjection to the combustor. With recycle and abovebed feed, SO_2 removal efficiency improved from less than 70 percent up to 94 percent, at 843°C (1500°F) and illustrates the impact that recycle has over the range of SO_2 control efficiencies under consideration in this report. With inbed feed and no recycle, SO_2 removal dropped from 90 to 83 percent over the temperature range of 793° to 871°C (1460° to 1600°F). SO_2 removal efficiency improved with recycle up to a level of about 94 percent, the same as measured with above-bed feed and recycle.

These results illustrate that above-bed feed of coal and limestone is appropriate for efficient SO_2 control as long as primary recycle is used. Since abovebed feeding may be simpler and less expensive than inbed feeding, these results set a favorable precedent in lowering FBC system cost. (This provides support to our contention in Section 4.0 that the cost of "best systems" of SO_2 control using FBC can be estimated by assuming abovebed feed with primary recycle (see Section 4.0)).

7.5.7 FluiDyne 3.3 ft × 5.3 ft Vertical Slice Combustor

The results of two runs are presented here, Run 35, and the 500-hr test run. The testing was done with Owatonna dolomite in both cases, and high gas phase residence times (>0.85 sec).

In the 500-hr test (begun September 20, 1977), the objective was to reduce SO_2 emissions to below 516 ng/J (1.2 lb/10^6 Btu), or a control efficiency of about 80 percent. Therefore, the results should not be interpreted as the most efficient control possible. Required Ca/S ratios ranged from 1.7 at 796°C (1465°F) to 2.4 at 718°C (1325°F), both with primary recycle. Although the gas residence time was longer for the testing at 718°C (1325°F), 1.5 versus 1.0 sec, and the excess air was much higher, 130 percent versus 30 percent, the Ca/S requirement was probably greater because of the low temperature and inefficient calcining of the available $CaCO_3$. The effect of excess air at 130 percent is uncertain, but it may have allowed for better SO_2 capture than would have been attained at 718°C (1325°F) if a lower excess air rate were used.

The dolomite particle size was the same at both temperatures, 6350 μm × 0 (1/4 in. × 0). Although the average size is not known, it is likely that it was greater than 500 μm. If so, one could speculate that even better performance could have been attained at smaller particles sizes.

Run 35 was performed with above-bed feed and recycle using dolomite (6350 μm × 0) and a gas phase residence time of 0.86 sec. An SO_2 removal efficiency of 87.2 percent was attained at a Ca/S ratio of 2.38. This lends further support to the ability of FBC to perform efficiently with above-bed feed and primary recycle.

7.5.8 National Coal Board 6-in. Diameter Unit

The results of this testing are itemized in Tables 88 through 90, and summarized in Table 92. Of the three criteria pollutants, only SO_2 data were reported. In one series of runs, U.K. limestone was used during combustion of Welbeck, Park Hill, Illinois, and Pittsburgh coals. Fluidizing velocity varied between 0.6 to 0.9 m/sec (2 to 3 ft/sec) but in most cases the unit was operated at 0.9 m/sec (3 ft/sec), so that gas phase residence time was generally 0.67 sec, based on an expanded bed depth of 0.6 m (2 ft). NCB forwarded two possible explanations to account for the better results obtained during Welbeck coal combustion. First, NCB found that the total rate of sulfur release from Welbeck coal was more rapid than for Pittsburgh coal (the other two coals were not tested). This may have minimized the quantity of sulfur released from elutriated fines in the freeboard, where reaction with sorbent is inefficient A second explanation was that because of the low feed rate of sorbent with low sulfur Welbeck coal, the bed residence time of coarse sorbent particles may have been longer. SO_2 emission control performance was excellent regardless of coal type in this set of experiments. Except for one experimental case, 90 percent SO_2 removal was achieved at a Ca/S molar feed ratio of 3 or less. This is not surprising since the actual operating conditions corresponded closely with "best system" operating conditions. U.K. limestone was prepared to a median particle diameter of 537 μm so that average in-bed particle size was probably close to 500 μm or slightly less.

Another set of experiments was run with limestone 1359 and Illinois coal. SO_2 reduction was improved when bed depth was expanded to 0.9 m (3 ft) from 0.6 m (2 ft), as would be expected. Use of finely crushed (-125 μm) limestone also improved performance, although primary recycle is absolutely essential in this operating mode to control the high sorbent elutriation rate. The overall results indicate that limestone 1359 was less effective than U.K. limestone in controlling SO_2 emissions. This result is expected since the reactivity of limestone 1359 is less than average.

A final set of experiments was reported for the NCB 6-in. test unit using limestone 18 with Pittsburgh (five tests) and Welbeck (one test) coals. Limestone 18 proved more effective with Pittsburgh coal than did U.K. limestone. The one test with Welbeck coal indicated performance similar to testing with U.K. limestone. The major difference in this series of tests was that limestone was finely crushed to a median size of 207 µm.

SO_2 removal performance was generally good in all three sets of experiments. This results from the proximity of operating conditions to recommended "best" operating conditions.

7.5.9 Argonne National Laboratory (ANL)

The results of testing on the ANL 6-in. unit are tabulated in Table 96 and summarized in Tables 92 and 93. SO_2 and NO_x data are reported.

Although the unit is small and the data was generated between 1970 and 1973, it is quite comprehensive and still useful.

The data demonstrates the ability of FBC to operate at the "best system" conditions and achieve very good SO_2 reduction results with reasonably low Ca/S ratios. The information also illustrates that for the same unit using the same Ca/S ratios, the reduction efficiency can vary widely with relation to the gas phase residence time.

The NO_x data on the other hand is not as representative of the actual values expected from larger units. The values appear considerably higher than data from the B&W 6 ft × 6 ft unit and the Renfrew unit (the two largest units for which data is reported). Even so, more than two-thirds of the data listed is below 301 ng/J (0.7 lb/10^6 Btu), the moderate level of control.

The majority of the tests were run with gas residence times between 0.66 and 1.00 sec. Two of the test series were run at 0.22 and 0.33 sec. Temperatures ranged from 718° to 900°C (1325° to 1650°F). Most tests were run using limestone 1359 with relatively small average particle sizes. Variations in sorbent included, dolomite 1337, limestone 1360, Tymochtee dolomite and a British sorbent referred to as B-Sonk. Ca/S ratios varied from 0 to 5.1 with the majority between 1.5 and 3.0. The figures in Subsection 7.6 show some of the ANL data used to extrapolate necessary Ca/S ratios for the 75, 85, and 90 percent control levels at close to "best system" conditions.

7.6 DERIVATION OF Ca/S RATIOS PRESENTED IN SECTION 3.0 FOR "BEST SYSTEM" OF SO_2 EMISSION REDUCTION

The Ca/S ratios presented in Table 22 in Section 3.0 were estimated by GCA from summary graphs of SO_2 reduction data which has been presented in tabulated form. The graphs are shown in this subsection, and are based on experimental results obtained from test units operated at or near "best system" conditions. A tabulation of important operating parameters is inset into each graph along with the interpolated Ca/S ratios at the optional SO_2 control levels. An index of graphs is listed below:

A. Figure 78 – Argonne National Laboratory, 6-in. diameter test unit using limestone 1359, 25 μm average particle size.

B. Figure 79 – Argonne National Laboratory, 6-in. diameter test unit using limestone 1359, 177 μm × 0 particle size distribution.

C. Figure 80 - National Coal Board, 36 in. × 18 in. diameter combustor using limestone 18, 1680 μm × 0 particle size distribution.

D. Figure 81 - Argonne National Laboratory, 6-in. diameter test unit using calcined limestone 1359, 25 μm average particle size.

E. Figure 82 - Argonne National Laboratory, 6-in. diameter test unit using limestone 1359, 490 to 630 μm average particle size.

F. Figure 83 - National Coal Board, 36 in. × 18 in. combustor using dolomite 1337, 1680 × 0 μm particle size distribution.

G. Figure 84 - National Coal Board, 36 in. × 18 in. combustor using limestone 18, 1680 × 0 μm particle size distribution.

H. Figure 85 - National Coal Board, 6-in. diameter test unit using U.K. limestone, 125 μm × 0 particle size distribution.

I. Figure 86 - National Coal Board, 6-in. diameter test unit using limestone 1359, 1680 μm × 0 and 125 μm × 0 particle size distribution.

J. Figure 87 - National Coal Board, 6-in. diameter and 36 in. 18 in. combustor using limestone 18, 1680 μm particle size distribution.

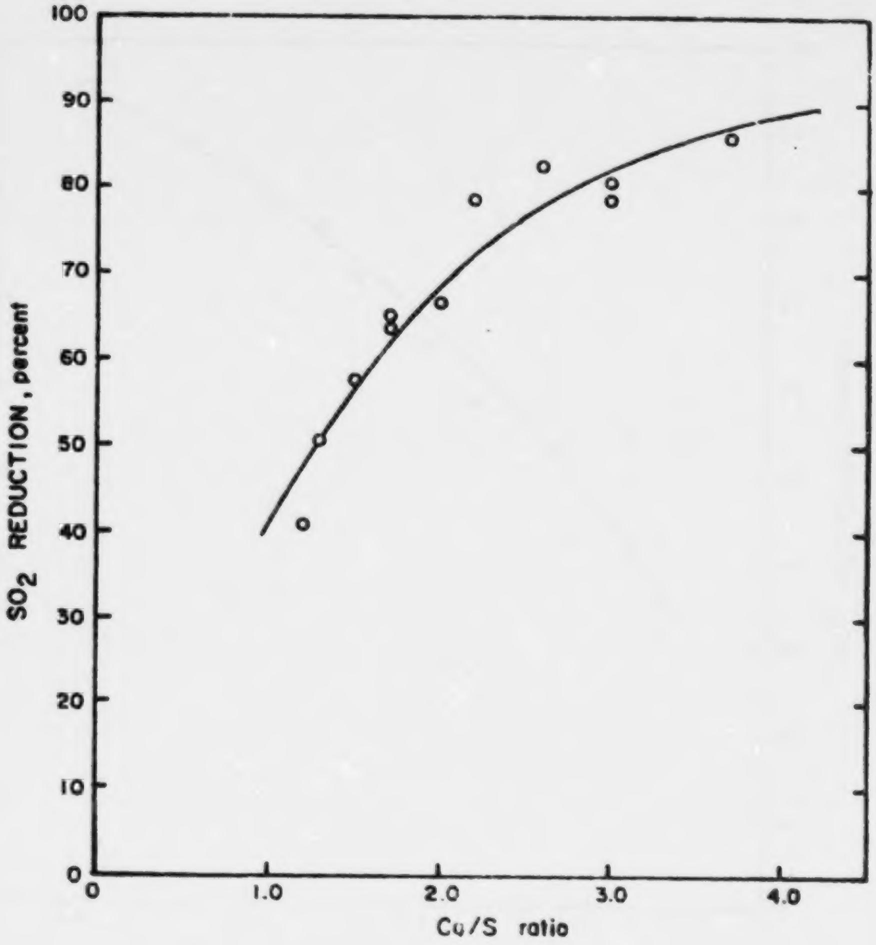
K. Figure 88 - Argonne National Laboratory and National Coal Board, 6-in. diameter test units using U.K. limestone.

L. Figure 89 - Argonne National Laboratory and National Coal Board, 6-in. diameter test units using limestone 1359.

7.7 COMPARISON OF EXPERIMENTAL DATA WITH WESTINGHOUSE SO_2 REMOVAL KINETIC MODEL

7.7.1 Westinghouse Studies

Westinghouse has compared experimental FBC SO_2 removal measurements with their projections of Ca/S requirements to confirm the SO_2 removal model. They concluded from their computerized file of FBC data that thermogravimetric projections are representative for the limited bench scale and pilot plant



INVESTIGATOR: ARGONNE NATIONAL LABORATORY

FBC UNIT DESIGNATION: 6" DIAMETER BENCH SCALE

SORBENT: LIMESTONE 1359

APPROXIMATE REACTIVITY: LOW

PARTICLE SIZE: 25μm AVERAGE

TEMPERATURE, °C (°F): 843-871 (1550-1600)

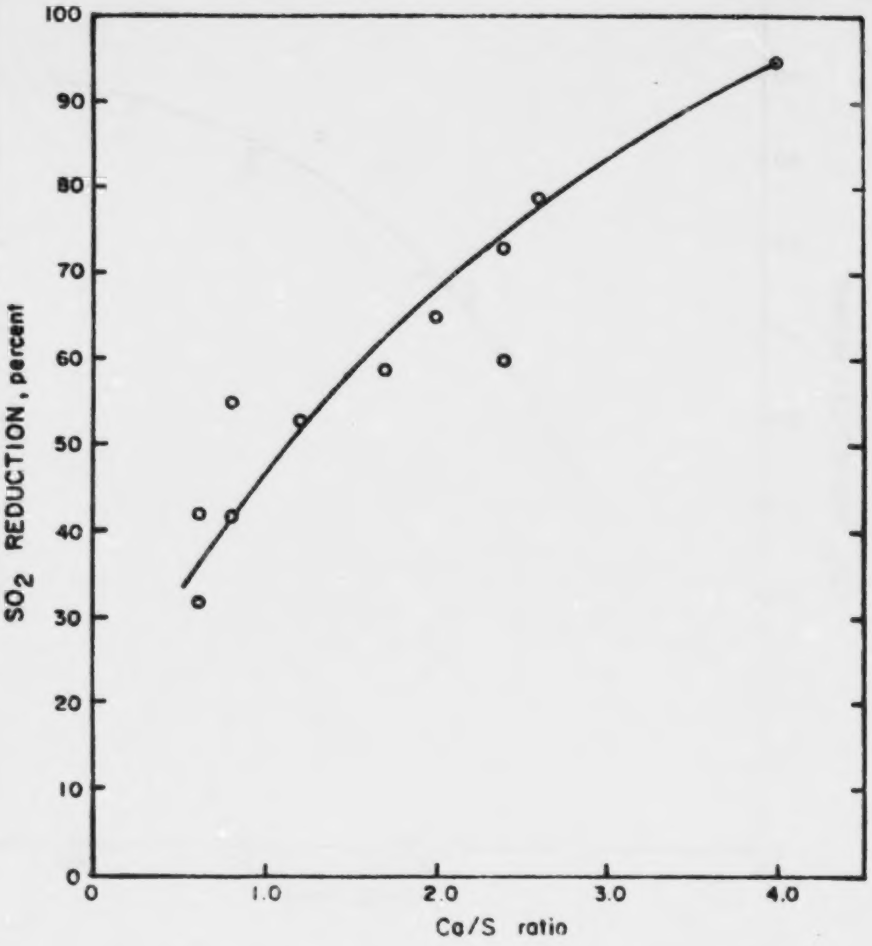
GAS PHASE

RESIDENCE TIME, seconds: 0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA:

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	2.4	2.7	3.1	3.2	3.4	4.2

Figure 78. Argonne National Laboratory, 6-in. diameter test unit using limestone 1359, 25 μm average particle size.



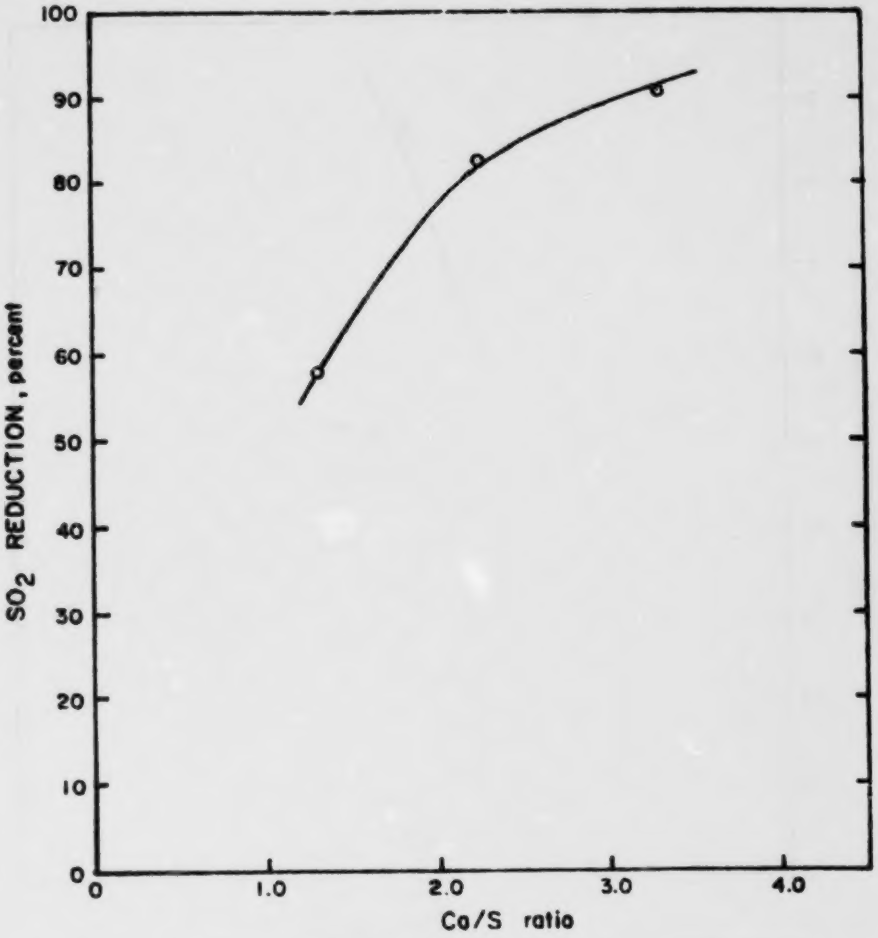
INVESTIGATOR: ARGONNE NATIONAL LABORATORY

FBC UNIT DESIGNATION: 6" DIAMETER BENCH SCALE
SORBENT: LIMESTONE 1359
APPROXIMATE REACTIVITY: LOW
PARTICLE SIZE: 177 μm x 0
TEMPERATURE, °C (°F): 843-871 (1550-1600)
GAS PHASE
RESIDENCE TIME, seconds: 0.67-0.7

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA:

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	2.5	2.7	2.9	3.0	3.1	3.6

Figure 79. Argonne National Laboratory, 6-in. diameter test unit using limestone 1359, 177 μm × 0 particle size distribution.



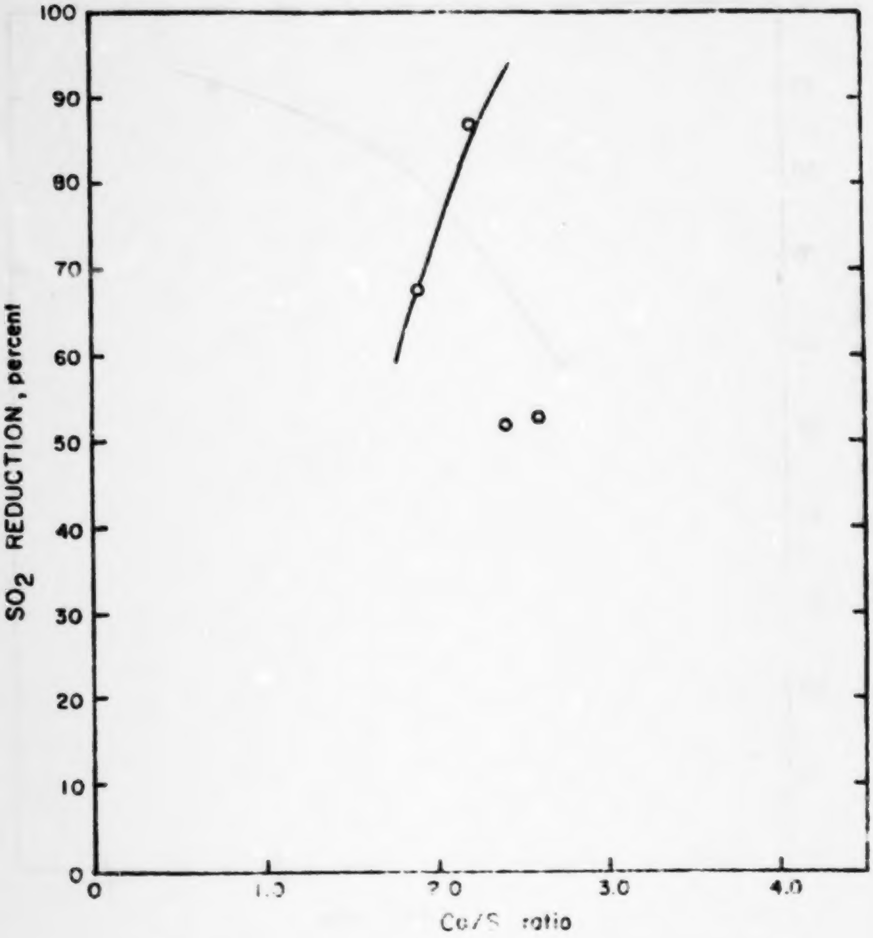
INVESTIGATOR: NATIONAL COAL BOARD

FBC UNIT DESIGNATION: 36" x 18" COMBUSTOR
SORBENT: LIMESTONE 18
APPROXIMATE REACTIVITY: HIGH
PARTICLE SIZE: 1680 μm x 0
TEMPERATURE, °C (°F): 849 (1560)
GAS PHASE
RESIDENCE TIME, seconds: 0.58

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA:

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	1.9	2.0	2.3	2.4	2.5	3.1

Figure 80. National Coal Board, 36 in. × 18 in. diameter combustor using limestone 18, 1680 μm × 0 particle size distribution.



INVESTIGATOR: ARGONNE NATIONAL LABORATORY

FBC UNIT DESIGNATION: 6" DIAMETER BENCH SCALE

SORBENT: CALCINED LIMESTONE 1359

APPROXIMATE REACTIVITY: HIGH

PARTICLE SIZE: 25 μm AVERAGE

TEMPERATURE, °C (°F): 871 (1600)

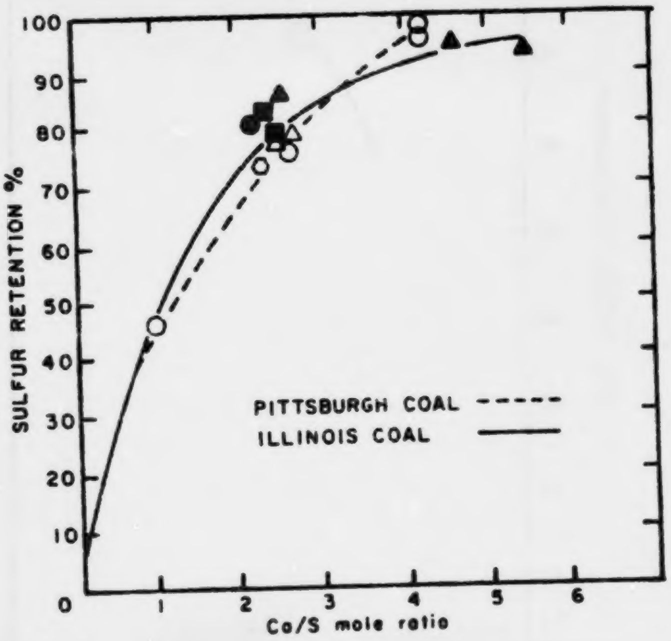
GAS PHASE

RESIDENCE TIME, seconds: 0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA:

% REMOVAL	75	78.7	63.2	83.9	85	90
Ca/S ratio,	2.0	2.0	2.1	2.1	2.2	2.3

Figure 81. Argonne National Laboratory, 6-in. diameter test unit using calcined limestone 1359, 25 μm average particle size.



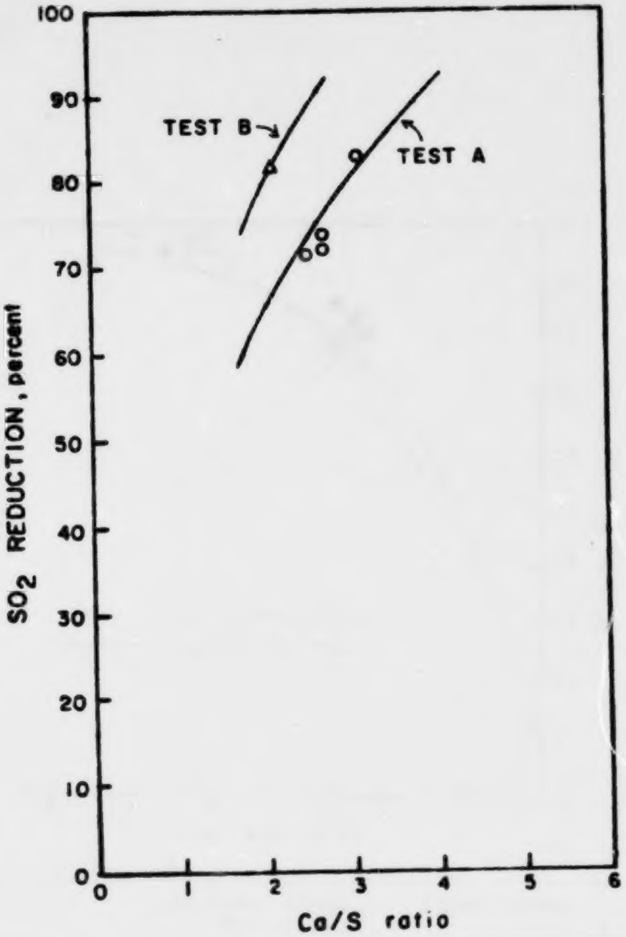
INVESTIGATOR: ARGONNE NATIONAL LABORATORY

FBC UNIT DESIGNATION: 6" DIAMETER BENCH SCALE
SORBENT: LIMESTONE 1359
APPROXIMATE REACTIVITY: LOW
PARTICLE SIZE: 490 - 630 μm AVERAGE
TEMPERATURE, °C (°F): 871 (1600)
GAS PHASE
RESIDENCE TIME, seconds: 0.5 - 0.7

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	2.1	2.5	2.7	2.8	3.0	3.5

Figure 82. Argonne National Laboratory, 6-in. diameter test unit using limestone 1359, 490 to 630 μm average particle size.



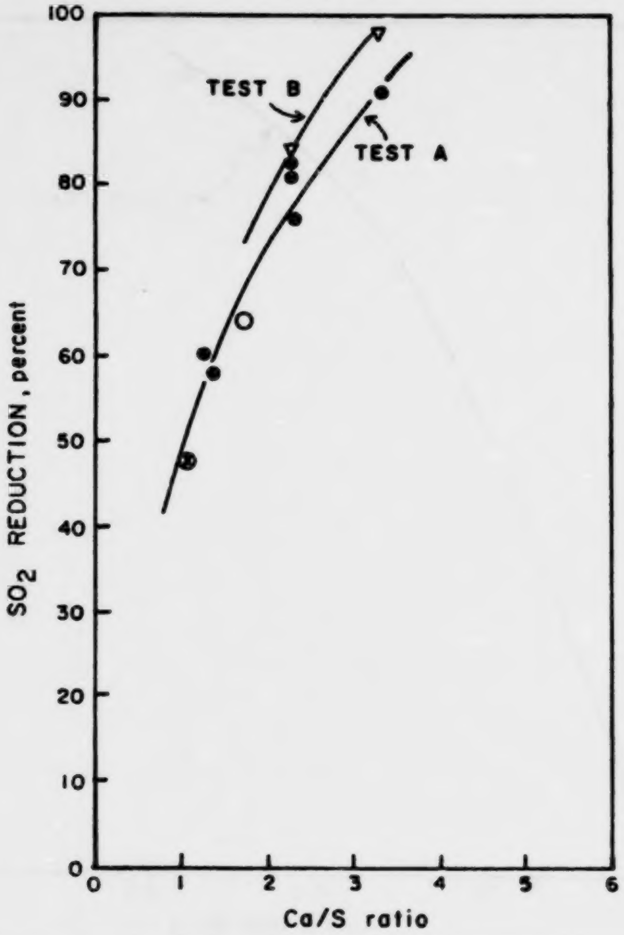
INVESTIGATOR: NATIONAL COAL BOARD

TEST:	A	B
FBC UNIT DESIGNATION:	36" x 18" COMBUSTOR	36" x 18" COMBUSTOR
SORBENT:	DOLOMITE 1337	DOLOMITE 1337
APPROXIMATE REACTIVITY:	HIGH	HIGH
PARTICLE SIZE:	1680 μm x 0	1680 μm x 0
TEMPERATURE, °C (°F):	799-849 (1470-1560)	749-849 (1380-1560)
GAS PHASE		
RESIDENCE TIME, seconds:	0.5	1.86

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,						
TEST A	2.6	3.0	3.3	3.3	3.4	3.8
TEST B	1.8	1.9	2.2	2.3	2.3	2.6

Figure 83. National Coal Board, 36 in. × 18 in. combustor using dolomite 1337, 1680 × 0 μm particle size distribution.

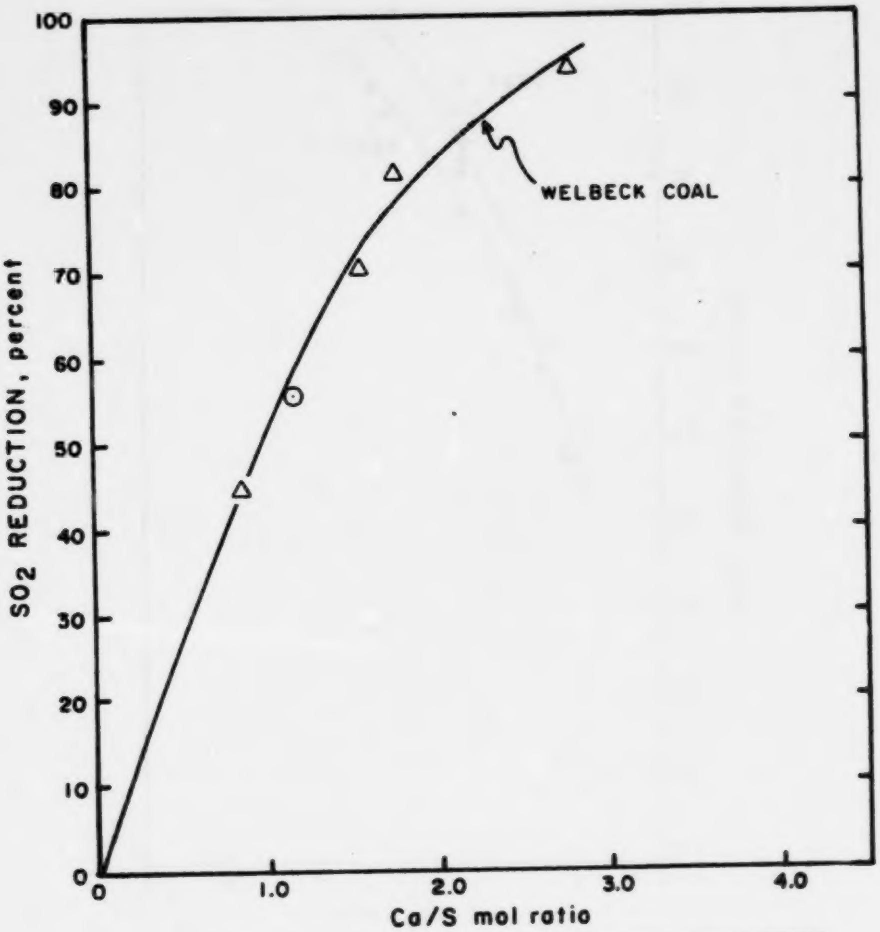


INVESTIGATOR: NATIONAL COAL BOARD		
TEST:	A	B
FBC UNIT DESIGNATION:	36" DIAMETER COMBUSTOR	36" DIAMETER COMBUSTOR
SORBENT:	LIMESTONE 18	LIMESTONE 18
APPROXIMATE REACTIVITY:	HIGH	HIGH
PARTICLE SIZE:	1680 μm x 0	1680 μm x 0
TEMPERATURE, °C (°F):	799-849 (1470-1560)	799-849 (1470-1560)
GAS PHASE RESIDENCE TIME, seconds:	0.5	0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,						
TEST A	2.1	2.3	2.7	2.7	2.8	3.2
TEST B	1.8	1.9	2.2	2.3	2.3	2.6

Figure 84. National Coal Board, 36 in. × 18 in. combustor using limestone 18, 1680 × 0 μm particle size distribution.



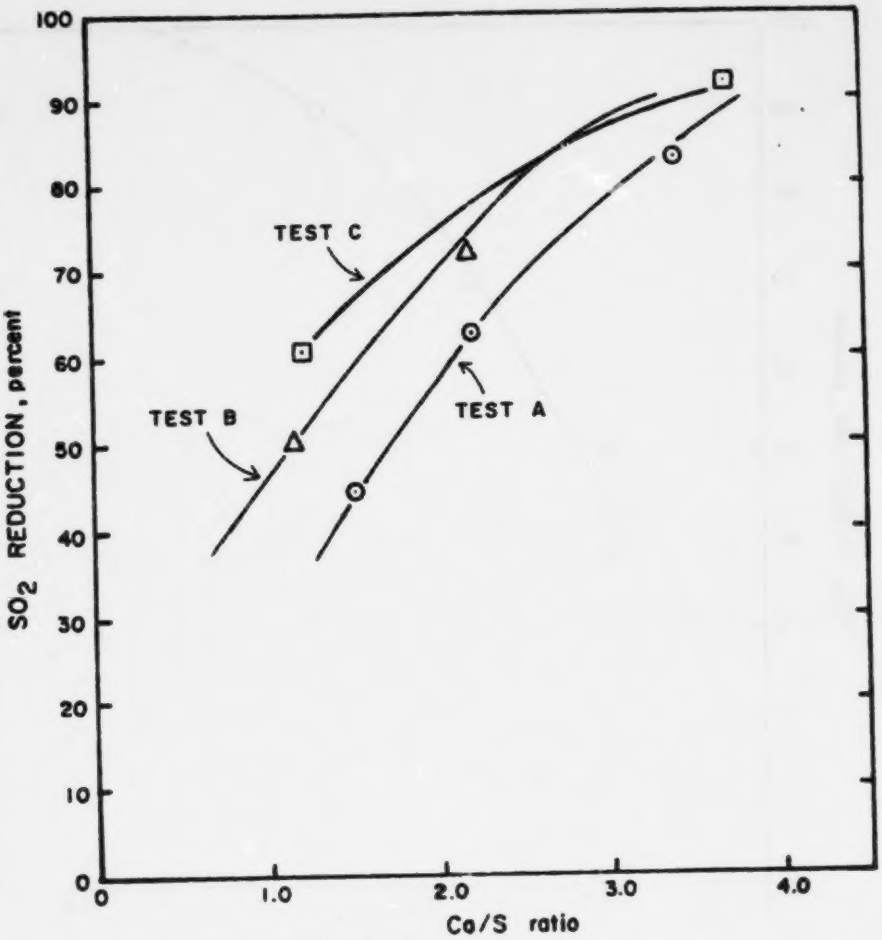
INVESTIGATOR: NATIONAL COAL BOARD

FBC UNIT DESIGNATION: 6" COMBUSTOR
SORBENT: U.K. LIMESTONE
APPROXIMATE REACTIVITY: HIGH
PARTICLE SIZE: 125 μm x 0
TEMPERATURE, °C (°F): 799 - (1470)
GAS PHASE
RESIDENCE TIME, seconds: 0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	1.6	1.8	2.0	2.0	2.1	2.4

Figure 85. National Coal Board, 6-in. diameter test unit using U.K. limestone, 125 μm × 0 particle size distribution.

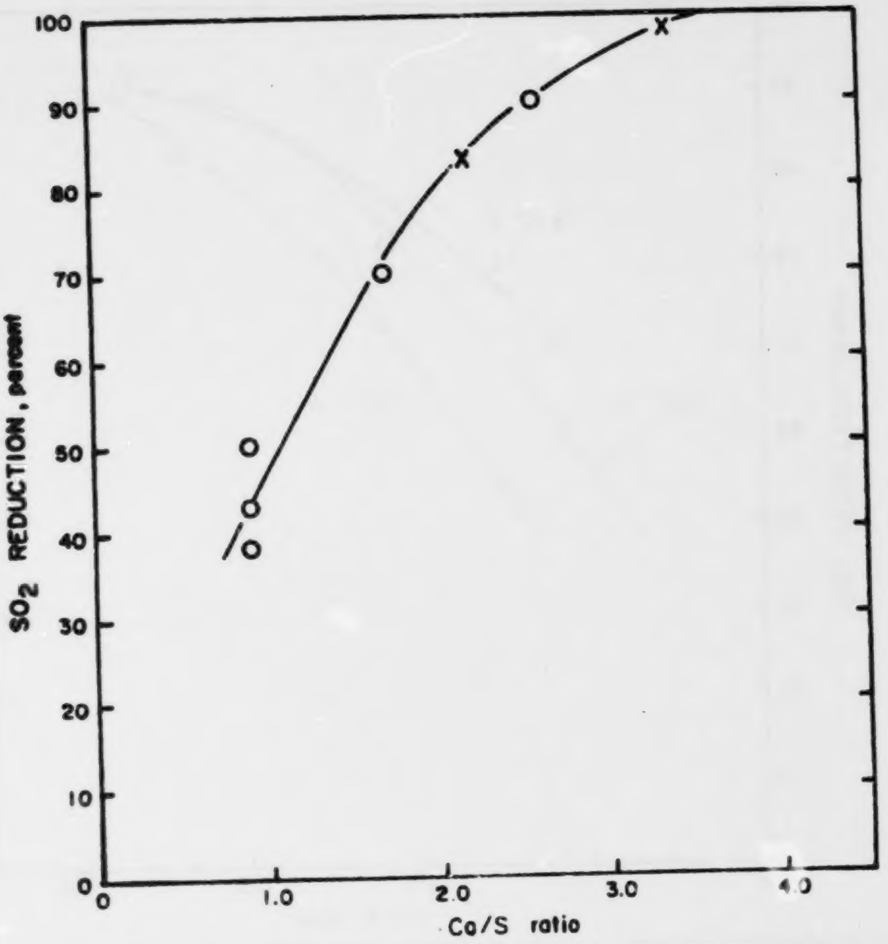


INVESTIGATOR: NATIONAL COAL BOARD			
TEST:	A	B	C
FBC UNIT DESIGNATION:	6" DIAMETER COMBUSTOR	6" DIAMETER COMBUSTOR	6" DIAMETER COMBUSTOR
SORBENT:	LIMESTONE 1359	LIMESTONE 1359	LIMESTONE 1359
APPROXIMATE REACTIVITY:	LOW	LOW	LOW
PARTICLE SIZE:	1680 μm x 0	1680 μm x 0	125 μm x 0
TEMPERATURE, °C (°F):	799 (1470)	799 (1470)	799 (1470)
GAS PHASE RESIDENCE TIME, seconds:	0.67	1.00	0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,						
TEST A	2.8	3.0	3.4	3.5	3.5	3.8
TEST B	2.3	2.4	2.7	2.7	2.8	3.3
TEST C	2.0	2.3	2.7	2.7	2.8	3.5

Figure 86. National Coal Board, 6-in. diameter test unit using limestone 1359, 1680 μm × 0 and 125 μm × 0 particle size distribution.



INVESTIGATOR: NATIONAL COAL BOARD

FBC UNIT DESIGNATION: 6" AND 36" COMBUSTORS

SORBENT: LIMESTONE 18

APPROXIMATE REACTIVITY: HIGH

PARTICLE SIZE: 1680 μm x 0

TEMPERATURE, °C (°F): 799 (1470)

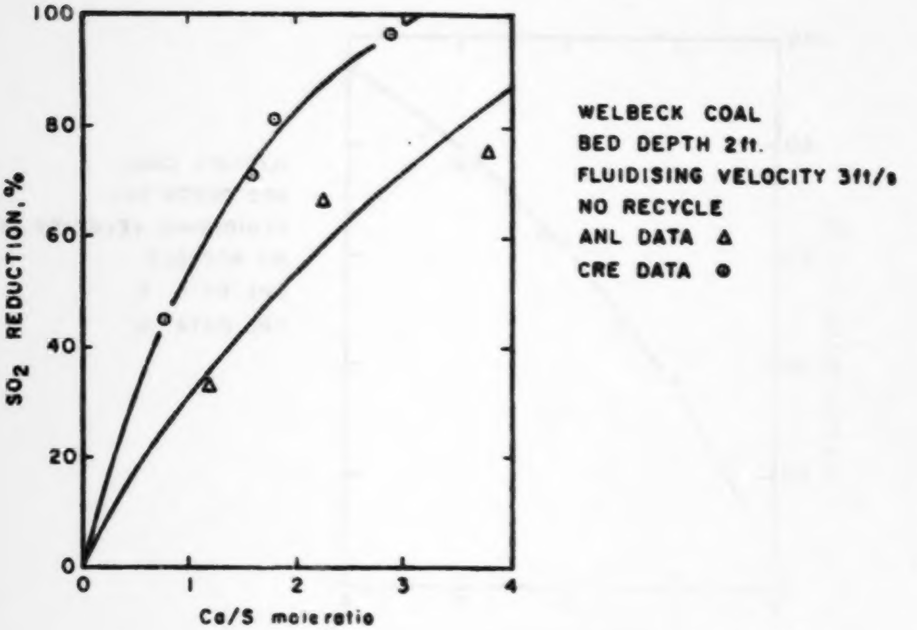
GAS PHASE

RESIDENCE TIME, seconds: 0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	1.8	1.9	2.1	2.2	2.2	2.6

Figure 87. National Coal Board, 6-in. diameter and 36 in. × 18 in. combustor using limestone 18, 1680 μm particle size distribution.



INVESTIGATOR: ARGONNE NATIONAL LABORATORY

FBC UNIT DESIGNATION: 6" UNITS

SORBENT: U.K. LIMESTONE

APPROXIMATE REACTIVITY:

PARTICLE SIZE: NOT REPORTED

TEMPERATURE, °C (°F): 799 (1470)

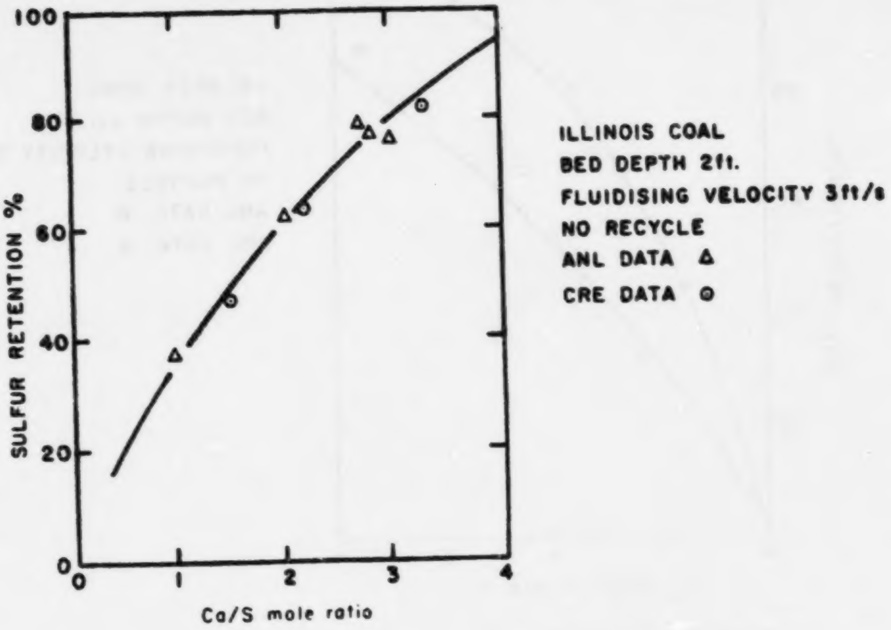
GAS PHASE

RESIDENCE TIME, seconds: 0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	3.2	3.4	3.6	3.7	3.8	4.2

Figure 88. Argonne National Laboratory and National Coal Board, 6-in. diameter test units using U.K. limestone.



INVESTIGATOR: ARGONNE NATIONAL LABORATORY

FBC UNIT DESIGNATION: 6" COMBUSTORS

SORBENT: LIMESTONE 1359

APPROXIMATE REACTIVITY: LOW

PARTICLE SIZE: NOT REPORTED

TEMPERATURE, °C (°F): 799 (1470)

GAS PHASE

RESIDENCE TIME, seconds: 0.67

SORBENT REQUIREMENTS FOR OPTIONAL SO_2 CONTROL LEVELS BASED ON THIS DATA

% REMOVAL	75	78.7	83.2	83.9	85	90
Ca/S ratio,	2.7	3.0	3.2	3.3	3.4	3.8

Figure 89. Argonne National Laboratory and National Coal Board, 6-in. diameter test units using limestone 1359.

data available at comparable operating conditions. Model projections of the Ca/S molar feed ratios required for various levels of SO_2 removal in AFBC, as a function of limestone type, are compared to the data collected from the ANL and British Coal Research fluidized-bed units for limestone 1359 in Figure 90. Conditions for the fluidized bed experimental runs were:

- Pressure - kPa (1 atm)
- Sorbent type - limestone 1359
- Sorbent particle size - 490 to 630 μm in feed
- Superficial velocity - 0.8 to 0.85 m/sec (2.6 to 2.8 ft/sec)
- Temperature - 788° to 798°C (1450° to 1468°F)
- Bed height - 0.6 m (2 ft)
- Flue gas conditions - 3 percent O_2, 15 percent CO_2

The Westinghouse projections are based on thermogravimetric rate data from sulfation at 815°C (1500°F) in 0.5 percent SO_2, 4 percent O_2, and N_2. The sulfations were carried out with 420 to 500 μm particles of limestone, calcined at 815°C (1500°F) in 15 percent CO_2 and nitrogen. The gas residence time (based on input bed height and velocity) was 0.66 sec, as opposed to an experimental value of 0.74 sec used by ANL. This longer residence time may account for the slightly lower Ca/S molar feed ratio requirements in the ANL limestone 1359 data.

7.7.2 GCA Calculations Based on the Westinghouse Model

Projections of Ca/S molar feed ratio requirements for several levels of desulfurization have been calculated by GCA for comparison with experimental results from the following test units.

- B&W 6 ft × 6 ft (1978)
- B&W 3 ft × 3 ft (1976)
- NCB-CRE 6 in. (1969)
- PER-FBM 1.5 ft × 6 ft (1971)

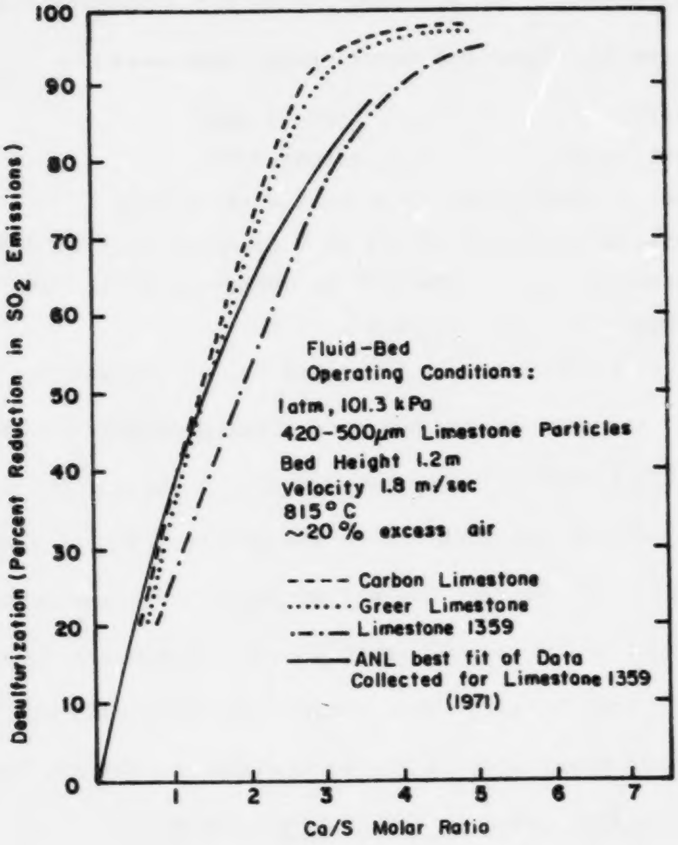


Figure 90. Comparison of experimental SO_2 data with projections based on Westinghouse Model.

Table 101 shows the sorbents used in the comparisons and the Westinghouse identification number for applicable thermogravimetric data. In general, the Westinghouse prediction of Ca/S requirements for various levels of sulfur removal are close to the experimental data measured. The comparisons are shown in Tables 97 through 100.

7.7.3 Influence of Fluidization Parameters Assumed in the Westinghouse Model

Important parametric values assumed for the projections are:

- ε - bed voidage = 0.5
- δ - volume fraction of bed bubbles = 0.5
- ρ_s - density of Ca in sorbent = 0.0271 mole Ca/cc
- $\bar{f}_h$ - fraction of bed volume occupied by heat transfer tubes = 0

The impact of the assumptions made for each of these parameters as used in the calculations are discussed below.

7.7.3.1 Particle Size Distribution--

The size of sorbent particles in the bed has a large effect on desulfurization efficiency. This is illustrated in the following example based on the B&W 3 ft × 3 ft operating conditions.

- Data Source: B&W 3 ft × 3 ft test no. 31 (see Table 82)

Operating conditions:		
	Temperature	- 819°C (1506°F)
	Bed height	- 0.4 m (16 in.)
	Superficial velocity	- 2.5 m/sec (8.1 ft/sec)
	SO_2 reduction	- 58 percent
	Ca/S ratio	- 2.46
	Sorbent type	- Lowellville
	Sorbent particle size	- 1000 μm × 0

Projections of sorbent needs using Westinghouse model based on carbon limestone (TG run number 231) are:

TABLE 96. SORBENTS USED EXPERIMENTALLY AND FOR PROJECTIONS USING WESTINGHOUSE MODEL

AFBC test unit	Experimental sorbent	Sorbent type used for Ca/S Projection	Westinghouse Identification No. for TG data	Bed particle size used for Ca/S calculation[*] (μm)	Sorbent particle size specified in experimental results, μm (feed size)
B&W 6 ft × 6 ft	Lowellville Limestone Lowellvile, Ohio	Carbon[†] Limestone	231	1,000	9,525 × 0 (average bed size - 1,600 μm)[‡]
B&W 3 ft × 3 ft	Greer Morgantown, W. Va.	Greer	86	1,000	2,380 × 0 (average bed size - 1,200 μm)[‡]
NCB-CRE 6 in.	Grove (Limestone 1359) Frederick, Md.	Grove	381	500	1,680 × 0 (average bed size - 400 μm)[§]
PER-FBM 1.5 ft × 6 ft	Grove (Limestone 1359)	Grove	296	75-150	44 × 0

[*] This assumed particle size was limited by the extent to which data was available from Westinghouse thermogravimetric experiments - 1,000 μm was the largest size reported in the Westinghouse experiments and 75 μm was the smaller size.

[†] Carbon limestone had the most similar sulfation characteristics based on the TG data available.

[‡] Based on size analysis of spent bed material.

[§] Assuming that average bed size is roughly one-half average feed size.

TABLE 97. COMPARISON OF EXPERIMENTAL AND PROJECTED SORBENT REQUIREMENTS FOR THE B&W 6 FT × 6 FT UNIT

Test No.	Bed temperature (^oC)	Gas residence time	Percent SO_2 removal	Required Ca/S ratios	
				Experimental	Projected
1-1	876	0.49	94.37	4.22	4.69
1-1	878	0.48	94.29	4.22	4.58
1-2	864	0.61	97.04	4.80	4.71
1-2	869	0.56	96.79	4.80	4.70
1-2	871	0.65	95.48	4.51	4.63
1-2	874	0.57	95.66	4.51	4.64
1-3	869	0.49	95.22	4.59	4.62
1-3	872	0.49	95.08	4.59	4.62
1-3	867	0.51	94.33	4.06	4.58
1-4	848	0.41	94.24	4.50	4.58
1-4	852	0.41	94.01	4.50	4.57
1-4	866	0.38	94.59	4.46	4.59
1-4	856	0.40	94.98	4.46	4.61
1-5	872	0.48	93.27	4.20	4.53

TABLE 98. COMPARISON OF EXPERIMENTAL AND PROJECTED SORBENT REQUIREMENTS FOR THE B&W 3 FT × 3 FT UNIT

Test No.	Bed temperature (^{o}C)	Gas residence time (seconds)	Percent SO_2 removal	Required Ca/S ratios	
				Experimental	Projected
46	837	0.18	81.7	3.62	3.14
47	838	0.16	85.0	3.94	3.27
48	843	0.14	48.3	2.70	2.20

TABLE 99. COMPARISON OF EXPERIMENTAL AND PROJECTED SORBENT REQUIREMENTS FOR THE PER FBM 1.5 FT × 6 FT UNIT

Test No.	Bed temperature (^{o}C)	Gas residence time	Percent SO_2 removal	Required Ca/S ratios	
				Experimental	Projected
27	854	0.21	74.0	2.0	2.15
28	871	0.21	71.6	2.4	2.06
	871	0.21	64.7	2.2	1.82
29	871	0.21	73.5	1.7	2.12
	871	0.21	73.5	2.0	2.12
30	882	0.20	50.0	1.4	1.37
	882	0.20	60.4	1.8	1.69
31	882	0.20	53.8	1.4	1.49
	882	0.20	61.5	1.8	1.72
32	877	0.20	61.9	1.6	1.73
	877	0.20	64.9	1.8	1.82
	877	0.20	70.5	1.8	2.02

TABLE 100. COMPARISON OF EXPERIMENTAL AND PROJECTED SORBENT REQUIREMENTS FOR THE NCB-CRE 6 IN. UNIT

Test No.	Bed temperature (^oC)	Gas residence time	Percent SO_2 removal	Required Ca/S ratios	
				Experimental	Projected
1-2	799	0.67	46.5	1.5	1.8
1-3	799	0.67	63.4	2.2	2.4
1-4	799	0.67	83.0	3.3	3.1
3-1	699	0.67	15.0	1.1	0.6
3-2	699	0.67	18.0	2.2	0.7
3-3	799	1.0	51.0	1.1	1.9
3-4	799	1.0	72.0	2.1	2.6
3-5	799	0.67	61.0	1.1	2.3
3-6	799	0.67	93.0	3.6	3.4

Average particle size considered	Projected Ca/S
1000 μm	3.87
500 μm	1.70
40 percent 500 μm 60 percent 1000 μm	2.56

The correlation between particle size and sorbent utilization exists because of the dependence of the sulfation reaction on mass transfer and inter- or intragranular diffusion. Mass transfer dominates only for about the first 10 percent of sulfation, but then diffusion becomes the rate limiting process. Diffusional resistance within the porous structure of the sorbent increases with particle size since sulfated outer regions limit diffusion into the interior of the particle.

7.7.3.2 Bed Voidage--

The gas residence time is an important consideration in achieving high efficiency SO_2 removal. Throughout this effort, it has been reported as the expanded bed height divided by the superficial velocity. However, for rigorous modeling purposes, correction factors are applied to determine interstitial velocity, which corrects for voidage, bed bubbles, and heat transfer tubes, as follows:

$$t = H/\mu$$

$$\mu = \frac{\mu_s}{(1 - \delta)\ \varepsilon + \delta\ \ (1 - \bar{f}_h)}$$

where t = gas residence time, sec
H = expanded bed depth
μ = interstitial gas velocity
μ_s = superficial gas velocity
δ = volume fraction of bed bubbles
ε = volume fraction of voids in a bed of particles
$\bar{f}_h$ = fraction of bed volume occupied by heat transfer surface

The following expression can be used to calculate the bed voidage if experimental data on static pressure as a function of elevation is given, as in the case of the B&W 3 ft × 3 ft unit.

$$\varepsilon = 1 - \frac{\Delta P/L}{\rho_s - \rho}$$

where ε = volume fraction of voids in a bed of particles

$\Delta P/L$ = pressure gradient

ρ_s = true particle density

ρ = fluid density

7.7.3.3 Bed Temperature--

A change in bed temperature has a strong effect on the sulfation rate constant of the sorbent because of the basic exponential dependence of the Arrhenius kinetics involved.

7.7.3.4 Solid Particle Density--

Uniform particle density throughout the particle distribution is important to provide a uniform fluidized system. This is assumed in applying the Westinghouse model.

7.8 EMISSION SOURCE TEST DATA FOR OIL-FIRED AFBC BOILERS

The only emission test data available in this category are results from the Argonne National Laboratory (ANL) 0.15 m (6-in.) bench scale experimental unit.[52] The size of the unit is small and nonrepresentative of expected commercial units. It is not warranted to present this data alone in support of emission standards development without other emissions data available from larger pilot and industrial scale units.

7.9 EMISSION SOURCE TEST DATA FOR GAS-FIRED AFBC BOILERS

There are no published emission source test data for gas-fired FBC boilers currently available.

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APPENDIX A

FIRST TIER OF AFBC COST ESTIMATES

The first tier (see Subsection 4.3) of AFBC Industrial Boiler Costs are included in this Appendix. The tabulated costs vary as a function of boiler capacity and coal type. The following costs are not included in the data in this Appendix.

- Capital costs
  - Limestone storage, conveying, and screening
  - Spent solids/ash conveying, and storage
- Operating costs
  - Limestone purchase
  - Spent solids/ash disposal
  - Electricity for operation of all auxiliary equipment

TABLE A-1. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS – 8.8 MW (30 x 10^6 Btu/h) EASTERN HIGH SULFUR COAL

Based on quote from	Company B	Capacity	8.8 MW (30 x 10^6 Btu/h)
Date of estimate	MID-1978	Coal Type	Eastern high sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	247,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	incl. w/ boiler
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	5,200
Boiler feed pumps*	13,400
Condensate systems*	7,700
Water treatment system*	8,000
Chemical feed*	1,400
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system (PEDCo – 20,000†)	116,700
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	399,400

*From PEDCo estimates for conventional systems.

†A cost of $20,000 for coal feeding equipment is included in the boiler cost.

TABLE A-1. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 35% x capital)	86,500
Stack	incl w/ boiler
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	2,500
Boiler feed pumps*	3,000
Condensate system*	1,100
Water treatment system*	2,000
Chemical feed*	800
Coal handling system (@ 60% x equipment)	70,000
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	31,600
Piping*	41,000
Insulation	incl. w/ boiler
Painting	5,900
Electrical	30,000
Buildings	140,400
TOTAL INSTALLATION COST	414,800
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	814,200
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	81,420
Construction & field expenses @ 10% DC	81,420
Construction fee @ 10% DC	81,420
Start-up and performance tests†	10,000
TOTAL INDIRECT COSTS (IC)	254,260
Contingencies @ 20% DC & IC	213,690
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	1,282,150
Land	2,000
Working capital @ 25% of total direct operating costs	120,300
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	1,404,450

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*From PEDCo estimates for conventional systems.

†Based on FBC vendor quotes.

TABLE A-1. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	157,900
Supervision*	68,500
Maintenance labor*	64,100
Replacement parts*	70,200
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	4,700
Fuel @$17./ton	113,480
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	2,300
TOTAL DIRECT COST	481,180
OVERHEAD	
Payroll (30% of direct labor)	47,370
Plant (26% of labor parts and maintenance)	93,770
TOTAL OVERHEAD COST	141,140
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	51,290
Capital recovery factor @10.6% Total turnkey cost	135,910
Interest on working capital @ 10% working capital	12,030
TOTAL CAPITAL CHARGES	199,230
TOTAL ANNUAL COSTS	821,550

*From PEDCo estimates for conventional systems.

TABLE A-2. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 8.8 MW (30 x 10^6 Btu/h) EASTERN LOW SULFUR COAL

Based on quote from Company B
Capacity 8.8MW(30x10^6Btu/h)
Date of estimate MID-1978
Coal Type Eastern low sulfur

CAPITAL EQUIPMENT COST

Boiler (with fans & ducts)	245,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	incl. w/ boiler
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	5,200
Boiler feed pumps*	13,400
Condensate systems*	7,700
Water treatment system*	8,000
Chemical feed*	1,400
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system (PEDCo - 20,000†)	96,800
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	377,500

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*From PEDCo estimates for conventional systems.

†A cost of $20,000 for coal feeding equipment is included in the boiler cost.

TABLE A-2. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 35% x capital)	86,000
Stack	incl. w/boiler
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	2,500
Boiler feed pumps*	3,000
Condensate system*	1,100
Water treatment system*	2,000
Chemical feed*	800
Coal handling system (@ 60% x equipment)	58,080
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	27,000
Piping*	35,000
Insulation	incl. w/ boiler
Painting	5,000
Electrical	30,000
Buildings	120,000
TOTAL INSTALLATION COST	370,480
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	747,980
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	74,800
Construction & field expenses @ 10% DC	74,800
Construction fee @ 10% DC	74,800
Start-up and performance tests†	10,000
TOTAL INDIRECT COSTS (IC)	234,400
Contingencies @ 20% DC & IC	196,480
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	1,178,860
Land	2,000
Working capital @ 25% of total direct operating costs	130,910
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	1,311,770

*From PEDCo estimates for conventional systems.

†Based on FBC vendor quotes.

TABLE A-2. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	157,900
Supervision*	68,500
Maintenance labor*	64,100
Replacement parts*	60,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	4,700
Fuel @ $29./ton	166,140
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	2,300
TOTAL DIRECT COST	523,640
OVERHEAD	
Payroll (30% of direct labor)	47,370
Plant (26% of labor parts and maintenance)	91,120
TOTAL OVERHEAD COST	138,490
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	47,150
Capital recovery factor @10.6% Total turnkey cost	124,960
Interest on working capital @ 10% working capital	13,090
TOTAL CAPITAL CHARGES	185,200
TOTAL ANNUAL COSTS	847,330

*From PEDCo estimates for conventional systems.

TABLE A-3. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 8.8 MW (30 x 10^6 Btu/h) SUBBITUMINOUS COAL

Based on quote from	Company B	Capacity	8.8 MW (30x10^6Btu/h)
Date of estimate	MID - 1978	Coal Type	Subbituminous

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	245,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	incl. w/ boiler
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	5,200
Boiler feed pumps*	13,400
Condensate systems*	7,700
Water treatment system*	8,000
Chemical feed*	1,400
Compressed air system (ID fan)	incl. w/ boiler
Coal handling system (PEDCo – 20,000†)	148,200
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	428,900

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*From PEDCo estimates for conventional systems.

†A cost of $20,000 for coal feeding equipment is included in the boiler cost.

TABLE A-3. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 35% x capital)	86,000
Stack	incl. w/ boiler
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/ coal handling
Deaerator*	2,500
Boiler feed pumps*	3,000
Condensate system*	1,100
Water treatment system*	2,000
Chemical feed*	800
Coal handling system (@ 60% x equipment)	88,920
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	38,900
Piping*	50,400
Insulation	incl. w/ boiler
Painting	7,200
Electrical	30,000
Buildings	172,800
TOTAL INSTALLATION COST	483,620
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	912,520
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	91,250
Construction & field expenses @ 10% DC	91,250
Construction fee @ 10% DC	91,250
Start-up and performance tests†	10,000
TOTAL INDIRECT COSTS (IC)	283,750
Contingencies @ 20% DC & IC	239,250
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	1,435,520
Land	2,000
Working capital @ 25% of total direct operating costs	109,810
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	1,547,330

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*From PEDCo estimates for conventional systems.

†Based on FBC vendor quotes.

TABLE A-3. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	157,900
Supervision*	68,500
Maintenance labor*	64,100
Replacement parts*	86,400
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	4,700
Fuel @$6.75/ton	55,350
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	2,300
TOTAL DIRECT COST	439,250
OVERHEAD	
Payroll (30% of direct labor)	47,370
Plant (26% of labor parts and maintenance)	97,980
TOTAL OVERHEAD COST	145,350
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	57,420
Capital recovery factor @10.6% Total turnkey cost	152,170
Interest on working capital @ 10% working capital	10,980
TOTAL CAPITAL CHARGES	220,570
TOTAL ANNUAL COSTS	805,170

*From PEDCo estimates for conventional systems.

TABLE A-4. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 22 MW (75 x 10^6 Btu/h) EASTERN HIGH SULFUR COAL

Based on quote from	Company A	Capacity	22MW(75 x 10^6 Btu/h)
Date of estimate	MID-1978	Coal Type	Eastern high sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	1,520,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	80,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	15,900
Boiler feed pumps*	25,300
Condensate systems*	8,700
Water treatment system*	15,000
Chemical feed*	1,400
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	165,400
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	1,831,700

*From PEDCo estimates for conventional systems.

TABLE A-4. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 35% x capital)	532,000
Stack	20,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	3,500
Boiler feed pumps*	5,500
Condensate system*	1,300
Water treatment system*	2,500
Chemical feed*	1,500
Coal handling system*	175,500
Spent solids withdrawal and cooling	incl. w/boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	84,200
Piping*	58,500
Insulation	incl. w/ boiler
Painting	8,200
Electrical	75,000
Buildings	234,000
TOTAL INSTALLATION COST	1,201,700
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	3,033,400
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	303,340
Construction & field expenses @ 10% DC	303,340
Construction fee @ 10% DC	303,340
Start-up and performance tests	67,670
TOTAL INDIRECT COSTS (IC)	977,690
Contingencies @ 20% DC & IC	802,220
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	4,813,310
Land	2,000
Working capital @ 25% of total direct operating costs	222,800
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	5,038,110

*From PEDCo estimates for conventional systems.

TABLE A-4. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	210,600
Supervision*	136,900
Maintenance labor*	128,200
Replacement parts*	117,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	9,500
Fuel @ $17. /ton	284,150
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	4,900
TOTAL DIRECT COST	891,250
OVERHEAD	
Payroll (30% of direct labor)	63,200
Plant (26% of labor parts and maintenance)	154,100
TOTAL OVERHEAD COST	217,300
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	192,500
Capital recovery factor @10.6% Total turnkey cost	510,200
Interest on working capital @ 10% working capital	22,300
TOTAL CAPITAL CHARGES	725,000
TOTAL ANNUAL COSTS	1,833,600

*From PEDCo estimates for conventional systems.

TABLE A-5. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 22 MW (75 x 10^6 Btu/h) EASTERN LOW SULFUR COAL

Based on quote from Company A

Date of estimate MID-1978

Capacity 22MW(75x10^6Btu/h)

Coal Type Eastern low sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	1,520,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	80,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	15,900
Boiler feed pumps*	25,300
Condensate systems*	8,700
Water treatment system*	15,000
Chemical feed*	1,400
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	141,400
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	1,807,700

*From PEDCo estimates for conventional systems.

TABLE A-5. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 35% x capital)	532,000
Stack	20,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	3,500
Boiler feed pumps*	5,500
Condensate system*	1,300
Water treatment system*	2,500
Chemical feed*	1,500
Coal handling system*	150,000
Spent solids withdrawal and cooling	incl. w/boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	72000
Piping*	53000
Insulation	incl. w/ boiler
Painting	7,000
Electrical	75,000
Buildings	200,000
TOTAL INSTALLATION COST	1,120,300
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	2,928,000
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	292,800
Construction & field expenses @ 10% DC	292,800
Construction fee @ 10% DC	292,800
Start-up and performance tests	65,560
TOTAL INDIRECT COSTS (IC)	943,960
Contingencies @ 20% DC & IC	774,400
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	4,646,360
Land	2,000
Working capital @ 25% of total direct operating costs	251,200
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	4,899,560

*From PEDCo estimates for conventional systems.

TABLE A-5. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	210,600
Supervision*	136,900
Maintenance labor*	128,200
Replacement parts*	100,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	9,500
Fuel @ $29 /ton	414,600
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	4,900
TOTAL DIRECT COST	1,004,700
OVERHEAD	
Payroll (30% of direct labor)	63,200
Plant (26% of labor parts and maintenance)	149,700
TOTAL OVERHEAD COST	212,900
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	185,900
Capital recovery factor @10.6% Total turnkey cost	492,500
Interest on working capital @ 10% working capital	25,100
TOTAL CAPITAL CHARGES	703,500
TOTAL ANNUAL COSTS	1,921,100

*From PEDCo estimates for conventional systems.

TABLE A-6. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 22 MW (30 x 10^6 Btu/h) SUBBITUMINOUS COAL

Based on quote from Company A
Date of estimate MID - 1978
Capacity 22 MW (75 x 10^6 Btu/h)
Coal Type Subbituminous

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	1,520,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	80,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	15,900
Boiler feed pumps*	25,300
Condensate systems*	8,700
Water treatment system*	15,000
Chemical feed*	1,400
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	203,600
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	1,869,900

*From PEDCo estimates for conventional systems.

TABLE A-6. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 35% x capital)	532,000
Stack	20,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	3,500
Boiler feed pumps*	5,500
Condensate system*	1,300
Water treatment system*	2,500
Chemical feed*	1,500
Coal handling system*	216,000
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	103,700
Piping*	72,000
Insulation	incl. w/ boiler
Painting	10,100
Electrical	75,000
Buildings	288,000
TOTAL INSTALLATION COST	1,331,100
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	3,201,000
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	320,100
Construction & field expenses @ 10% DC	320,100
Construction fee @ 10% DC	320,100
Start-up and performance tests	71,020
TOTAL INDIRECT COSTS (IC)	1,031,320
Contingencies @ 20% DC & IC	846,500
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	5,078,820
Land	2,000
Working capital @ 25% of total direct operating costs	193,200
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	5,274,020

*From PEDCo estimates for conventional systems.

TABLE A-6. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	210,600
Supervision*	136,900
Maintenance labor*	128,200
Replacement parts*	144,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	9,500
Fuel @$6.75/ton	138,720
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	4,900
TOTAL DIRECT COST	772,820
OVERHEAD	
Payroll (30% of direct labor)	63,200
Plant (26% of labor parts and maintenance)	161,100
TOTAL OVERHEAD COST	224,300
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	203,200
Capital recovery factor @10.6% Total turnkey cost	538,400
Interest on working capital @ 10% working capital	19,300
TOTAL CAPITAL CHARGES	760,900
TOTAL ANNUAL COSTS	1,758,020

*From PEDCo estimates for conventional systems.

TABLE A-7. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 44 MW (150 x 10^6 Btu/h) EASTERN HIGH SULFUR COAL

Based on quote from Company A — Capacity 44MW(150x10^6Btu/h)

Date of estimate MID-1978 — Coal Type Eastern high sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	2,427,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	300,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	21,600
Boiler feed pumps*	44,500
Condensate systems*	9,200
Water treatment system*	18,000
Chemical feed*	1,500
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	282,300
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	3,104,100

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*from PEDCo estimates for conventional systems

TABLE A-7. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 40% x capital)	970,000
Stack	50,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl w/coal handling
Deaerator*	4,000
Boiler feed pumps*	7,000
Condensate system*	1,500
Water treatment system*	3,000
Chemical feed*	1,500
Coal handling system*	292,500
Spent solids withdrawal and cooling	incl. w/boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	157,950
Piping*	81,900
Insulation	incl. w/ boiler
Painting	11,700
Electrical	150,000
Buildings	409,500
TOTAL INSTALLATION COST	2,141,350
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	5,245,450
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	524,550
Construction & field expenses @ 10% DC	524,550
Construction fee @ 10% DC	524,550
Start-up and performance tests	114,910
TOTAL INDIRECT COSTS (IC)	1,688,560
Contingencies @ 20% DC & IC	1,386,800
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	8,320,810
Land	2,000
Working capital @ 25% of total direct operating costs	352,030
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	8,674,840

*From PEDCo estimates for conventional systems.

TABLE A-7. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	315,900
Supervision*	136,900
Maintenance labor*	128,200
Replacement parts*	234,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	18,800
Fuel @$17 /ton	568,300
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	6,000
TOTAL DIRECT COST	1,408,100
OVERHEAD	
Payroll (30% of direct labor)	94,770
Plant (26% of labor parts and maintenance)	211,900
TOTAL OVERHEAD COST	306,670
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	332,800
Capital recovery factor @10.6% Total turnkey cost	882,000
Interest on working capital @ 10% working capital	35,200
TOTAL CAPITAL CHARGES	1,250,000
TOTAL ANNUAL COSTS	2,964,770

* from PEDCo estimates for conventional systems

TABLE A-8. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 44 MW (150 x 10^6 Btu/h) EASTERN LOW SULFUR COAL

Based on quote from Company A
Date of estimate MID - 1978
Capacity 44MW (150x10^6 Btu/h)
Coal Type Eastern low sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	2,427,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	300,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	21,600
Boiler feed pumps*	44,500
Condensate systems*	9,200
Water treatment system*	18,000
Chemical feed*	1,500
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	241,300
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	3,063,100

*from PEDCo estimates for conventional systems

TABLE A-8. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 40% x capital)	970,800
Stack	50,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/ coal handling
Deaerator*	4,000
Boiler feed pumps*	7,000
Condensate system*	1,500
Water treatment system*	3,000
Chemical feed*	1,500
Coal handling system*	250,000
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	135,500
Piping*	70,000
Insulation	incl. w/ boiler
Painting	10,000
Electrical	150,000
Buildings	350,000
TOTAL INSTALLATION COST	2,002,800
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	5,065,900
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	506,590
Construction & field expenses @ 10% DC	506,590
Construction fee @ 10% DC	506,590
Start-up and performance tests	111,320
TOTAL INDIRECT COSTS (IC)	1,631,090
Contingencies @ 20% DC & IC	1,339,400
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	8,036,400
Land	2,000
Working capital @ 25% of total direct operating costs	408,380
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	8,446,780

*From PEDCo estimates for conventional systems.

TABLE A-8. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	315,900
Supervision*	136,900
Maintenance labor*	128,200
Replacement parts*	200,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	18,800
Fuel @$29./ton	827,700
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	6,000
TOTAL DIRECT COST	1,633,500
OVERHEAD	
Payroll (30% of direct labor)	94770
Plant (26% of labor parts and maintenance)	203,070
TOTAL OVERHEAD COST	297,840
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	321,460
Capital recovery factor @10.6% Total turnkey cost	851,860
Interest on working capital @ 10% working capital	40,840
TOTAL CAPITAL CHARGES	1,214,160
TOTAL ANNUAL COSTS	3,145,500

*From PEDCo estimates for conventional systems.

TABLE A-9. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 44 MW (150 x 10^6 Btu/h) SUBBITUMINOUS COAL

Based on quote from Company A Capacity 44M(150 x10^6 Btu/h)

Date of estimate __________ Coal Type Subbituminous

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	2,427,000
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	300,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	21,600
Boiler feed pumps*	44,500
Condensate systems*	9,200
Water treatment system*	18,000
Chemical feed*	1,500
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	347,500
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	3,169,300

*From PEDCo estimates for conventional systems.

TABLE A-9. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 40% x capital)	970,200
Stack	50,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/ coal handling
Deaerator*	4,000
Boiler feed pumps*	7,000
Condensate system*	1,500
Water treatment system*	3,000
Chemical feed*	1,500
Coal handling system*	360,000
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	194,400
Piping*	100,800
Insulation	incl. w/ boiler
Painting	14,400
Electrical	150,000
Buildings	504,000
TOTAL INSTALLATION COST	2,361,400
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	5,530,700
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	553,070
Construction & field expenses @ 10% DC	553,070
Construction fee @ 10% DC	553,070
Start-up and performance tests	120,610
TOTAL INDIRECT COSTS (IC)	1,779,820
Contingencies @ 20% DC & IC	1,462,100
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	8,772,620
Land	2,000
Working capital @ 25% of total direct operating costs	292,730
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	9,067,350

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*From PEDCo estimates for conventional systems.

TABLE A-9. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	315,900
Supervision*	136,900
Maintenance labor*	128,200
Replacement parts*	288,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	18,800
Fuel @$6.75/ton	277,100
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	6,000
TOTAL DIRECT COST	1,170,900
OVERHEAD	
Payroll (30% of direct labor)	94,770
Plant (26% of labor parts and maintenance)	225,950
TOTAL OVERHEAD COST	320,720
By-product credits	N.A
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	350,900
Capital recovery factor @10.6% Total turnkey cost	929,900
Interest on working capital @ 10% working capital	29,270
TOTAL CAPITAL CHARGES	1,310,070
TOTAL ANNUAL COSTS	2,801,690

*From PEDCo estimates for conventional systems.

TABLE A-10. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 58.6 MW (200 x 10^6 Btu/h) EASTERN HIGH SULFUR COAL

Based on quote from Company A

Date of estimate MID - 1978

Capacity 58.6 MW (200 x 10^6 Btu/h)

Coal Type Eastern high sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	3,111,500
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	365,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	29,000
Boiler feed pumps*	58,000
Condensate systems*	16,000
Water treatment system*	20,000
Chemical feed*	1,500
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	308,800
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	3,910,100

*From PEDCo estimates for conventional systems.

TABLE A-10. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 40% x capital)	1,244,600
Stack	60,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl w/coal handling
Deaerator*	5,000
Boiler feed pumps*	8,000
Condensate system*	2,000
Water treatment system*	3,500
Chemical feed*	1,500
Coal handling system*	321,800
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	189,500
Piping*	93,600
Insulation	incl. w/ boiler
Painting	11,700
Electrical	160,000
Buildings	444,600
TOTAL INSTALLATION COST	2,545,800
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	6,455,900
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	645,590
Construction & field expenses @ 10% DC	645,590
Construction fee @ 10% DC	645,590
Start-up and performance tests	139,120
TOTAL INDIRECT COSTS (IC)	2,075,890
Contingencies @ 20% DC & IC	1,706,360
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	10,238,150
Land	2,000
Working capital @ 25% of total direct operating costs	458,080
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	10,698,230

*From PEDCo estimates for conventional systems.

TABLE A-10. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	421,200
Supervision*	136,900
Maintenance labor*	192,200
Replacement parts*	292,500
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	25,200
Fuel @ $17. /ton	756,800
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	7,500
TOTAL DIRECT COST	1,832,300
OVERHEAD	
Payroll (30% of direct labor)	126,360
Plant (26% of labor parts and maintenance)	271,130
TOTAL OVERHEAD COST	397,490
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	409,530
Capital recovery factor @10.6% Total turnkey cost	1,085,240
Interest on working capital @ 10% working capital	45,810
TOTAL CAPITAL CHARGES	1,540,580
TOTAL ANNUAL COSTS	3,770,370

*From PEDCo estimates for conventional systems.

TABLE A-11. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 58.6 MW (200 x 10^6 Btu/h) EASTERN LOW SULFUR COAL

Based on quote from Company A Capacity 58.6MW(200x10^6 Btu/h)
Date of estimate MID-1978 Coal Type Eastern low sulfur

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	3,111,500
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	365,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	29,000
Boiler feed pumps*	58,000
Condensate systems*	16,300
Water treatment system*	20,000
Chemical feed*	1,500
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	263,900
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	3,865,200

*From PEDCo estimates for conventional systems.

TABLE A-11. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 40% x capital)	1,244,600
Stack	60,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	5,000
Boiler feed pumps*	8,000
Condensate system*	2,000
Water treatment system*	3,500
Chemical feed*	1,500
Coal handling system*	275,000
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	162,000
Piping*	80,000
Insulation	incl. w/ boiler
Painting	10,000
Electrical	160,000
Buildings	380,000
TOTAL INSTALLATION COST	2,391,600
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	6,256,800
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	625,680
Construction & field expenses @ 10% DC	625,680
Construction fee @ 10% DC	625,680
Start-up and performance tests	135,140
TOTAL INDIRECT COSTS (IC)	2,012,180
Contingencies @ 20% DC & IC	1,653,800
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	9,922,780
Land	2,000
Working capital @ 25% of total direct operating costs	534,520
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	10,459,300

*From PEDCo estimates for conventional systems.

TABLE A-11. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	421,200
Supervision*	136,900
Maintenance labor*	192,200
Replacement parts*	250,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	25,200
Fuel @$29 /ton	1,105,080
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	7,500
TOTAL DIRECT COST	2,138,080
OVERHEAD	
Payroll (30% of direct labor)	126,360
Plant (26% of labor parts and maintenance)	260,080
TOTAL OVERHEAD COST	386,440
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	396,910
Capital recovery factor @10.6% Total turnkey cost	1,051,810
Interest on working capital @ 10% working capital	53,450
TOTAL CAPITAL CHARGES	1,502,170
TOTAL ANNUAL COSTS	4,026,690

*From PEDCo estimates for conventional systems.

TABLE A-12. ESTIMATED CAPITAL, OPERATING AND ANNUALIZED COSTS FOR AFBC INDUSTRIAL BOILERS — 58.6 MW (200 x 10 Btu/h) SUBBITUMINOUS COAL

Based on quote from Company A
Capacity 58.6MW(200x10^6 Btu/h)
Date of estimate MID-1978
Coal Type Subbituminous

CAPITAL EQUIPMENT COST	
Boiler (with fans & ducts)	3,111,500
Primary particulate control device	incl. w/ boiler
Final particulate control device	not included
Stack	365,000
Instrumentation	incl. w/ boiler
Pulverizers	
Coal	NA
Limestone	NA
Feeders	
Coal	incl. w/ boiler
Limestone	incl. w/ boiler
Crushers	
Coal	incl. w/ coal handling
Limestone	screening is included in limestone handling & storage. See Table C-20
Deaerator*	29,000
Boiler feed pumps*	58,000
Condensate systems*	16,300
Water treatment system*	20,000
Chemical feed*	1,500
Compressed air system (FD fan)	incl. w/ boiler
Coal handling system*	380,000
Limestone handling & storage system	See Table C-20
Spent solids withdrawal & cooling system	incl. w/ boiler
Spent solids and ash handling and storage system	See Table C-21
TOTAL EQUIPMENT COST	3,981,300

*From PEDCo estimates for conventional systems.

TABLE A-12. (CONT'd)

EQUIPMENT INSTALLATION COSTS, DIRECT	
Boiler (@ 40% x capital)	1,244,600
Stack	60,000
Instrumentation	incl. w/ boiler
Pulverizers	NA
Feeders	incl. w/ boiler
Crushers	incl. w/coal handling
Deaerator*	5,000
Boiler feed pumps*	8,000
Condensate system*	2,000
Water treatment system*	3,500
Chemical feed*	1,500
Coal handling system*	396,000
Spent solids withdrawal and cooling	incl. w/ boiler
Limestone handling and storage system	See Table C-20
Spent solids and ash handling and storage system	See Table C-21
Foundation and Supports (@ 90% PEDCo estimate)	259,200
Piping*	115,200
Insulation	incl. w/ boiler
Painting	11,700
Electrical	160,000
Buildings	547,200
TOTAL INSTALLATION COST	2,813,900
TOTAL DIRECT COST (EQUIPMENT & INSTALLATION)(DC)	6,795,200
EQUIPMENT INSTALLATION COST (INDIRECT)	
Engineering @ 10% DC	679,520
Construction & field expenses @ 10% DC	679,520
Construction fee @ 10% DC	679,520
Start-up and performance tests	145,900
TOTAL INDIRECT COSTS (IC)	2,184,460
Contingencies @ 20% DC & IC	1,795,930
TOTAL TURNKEY COST (DC + IC + CONTINGENCIES)	10,775,590
Land	2,000
Working capital @ 25% of total direct operating costs	378,180
GRAND TOTAL CAPITAL COST (TURNKEY + LAND + WORKING CAPITAL)	11,155,770

*From PEDCo estimates for conventional systems.

TABLE A-12. (CONT'd)

DIRECT OPERATING COST	
Direct labor*	421,200
Supervision*	136,900
Maintenance labor*	192,200
Replacement parts*	360,000
Electricity	See Table C-24
Steam	NA
Cooling water	NA
Process water*	25,200
Fuel @ $6.75/ton	369,700
Limestone	See Table C-22
Waste disposal	See Table C-23
Chemicals*	7,500
TOTAL DIRECT COST	1,512,700
OVERHEAD	
Payroll (30% of direct labor)	126,360
Plant (26% of labor parts and maintenance)	288,680
TOTAL OVERHEAD COST	415,040
By-product credits	N.A.
CAPITAL CHARGES	
G&A, taxes & insurance @ 4% Total turnkey cost	431,020
Capital recovery factor @10.6% Total turnkey cost	1,142,210
Interest on working capital @ 10% working capital	37,820
TOTAL CAPITAL CHARGES	1,611,050
TOTAL ANNUAL COSTS	3,538,790

*From PEDCo estimates for conventional systems.

APPENDIX B

COST BASIS USED IN OTHER INDUSTRIAL FBC BOILER COST ESTIMATES

EXXON - APPLICATION OF FLUIDIZED-BED TECHNOLOGY TO INDUSTRIAL BOILERS

This report estimated costs for "grass roots" FBC and conventional industrial boilers producing 100,000 lb/hr steam at 125 psig. The important assumptions used for FBC costs are shown below. A complete listing is included in Appendix A1 of the original Exxon report.

Capital Costs

- Two boilers for each case, each rated at 100,000 lb/hr steam and 82 percent efficiency.
- U.S. Gulf Coast Location, First Quarter, 1975
- Process development allowance of 15 percent added to FBC cost
- Environmental standards for coal firing:
  - SO_2 - 516 ng/J (1.2 lb/10^6 Btu)
  - NO_x - 301 ng/J (0.7 lb/10^6 Btu)
  - Particulate - 43 ng/J (0.1 lb/10^6 Btu)
- Coal: Illinois No. 6; 3.6 percent S, 8.0 percent ash, HHV = 10,600 Btu/lb
- Coal and limestone handling:
  - Coal - 10 day storage, ready for charging as delivered
  - Limestone - 10 day storage, 1/8 in. particle size
- Solid waste handling: stored and hauled to disposal by truck.

Items excluded from the capital estimate are:

- Land;
- Unusual site preparation;
- Boiler feedwater treatment facilities (included as operating costs);
- Blowdown system; and
- Steam distribution system

Operating costs were derived using the following basis:

- Load factor = 0.9
- Manpower = 20,000 $/yr/man
- Electricity = 4¢/kwh
- Limestone = 12$/ton
- Waste solids disposal = 8$/ton
- Annual repair materials = 1.5 percent of investment
- Annual cost for supplies, local taxes, administrative expense, and general expense = 3.0 percent of investment
- Annual capital charges = 20 percent of investment
- Boiler feedwater and blowdown cost = 60¢/1,000 lb of produced steam

Adjustments made to Exxon estimates to achieve comparability with cost estimates derived here, are shown in Table B-1. Only the high sulfur coal case was considered.

These adjustments result in an annualized capital charge of $\$2.30/10^6$ Btu output and a total annual cost of $\$6.14/10^6$ Btu output.

A.G.McKEE - 100,000 LB/HR BOILER COST STUDY

The McKee study considers three boiler systems rated at 100,000 lb/hr steam, including:

TABLE B-1. ADJUSTMENTS MADE TO EXXON COST BASIS

1. The cost of one boiler and total ESP cost were subtracted.
2. The process development allowance of 20 percent was subtracted.
3. A load factor of 0.6 (as opposed to 0.9) was used to determine annualized capital cost.
4. The Marshall Stevens equipment index for steam power was used to update capital costs from First Quarter, 1975 to Third Quarter, 1978.
5. A cost of $\$0.88/10^6$ Btu output was used for Eastern high sulfur coal based on 82 percent boiler efficiency and \$17/ton of coal.
6. Operating costs were updated by a factor of 7 percent/yr.

- AFBC burning noncompliance coal;
- Conventional spreader stoker burning noncompliance coal with a mechanical collector and double alkali FGD; and
- Conventional spreader stoker-burning compliance coal with dry ESP.

The AFBC costs are based on the current 1978 contract costs associated with installation of a boiler at Georgetown University in Washington, D.C. Therefore, a minimum amount of cost estimating was required for the AFBC case. The costs for the comparable conventional boilers were based on McKee's own inhouse data. The equipment included in the AFBC system includes:

- One, 100,000 lb/hr-steam AFBC boiler top supported operating at 625 psig saturated steam consisting of several shop-assembled components including lagging, insulation, and setting.
- Coal receiving, conveying system, crushing, screening, storage, weighing and spreader feeder system.
- Solid waste material cooling, conveying, storage and disposal system. (Two waste materials - bottom ash and top ash.)*
- Combustion air supply system.
- Flue gas exhaust system.
- Mechanical collector and reinjection system.
- Economizer.
- Bag filter and disposal system.
- Fuel oil startup system with flame safety.

Comparable equipment was included in the estimates of conventional boiler cost. The following equipment was not included in any of the systems:

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*In this particular FBC system bottom ash or spent bed material is drained from the fluid bed continuously to maintain a constant bed level. The bottom ash is cooled, crushed, stored and hauled separately because of its potential value as a chemical. Top ash consists primarily of coal ash and is removed from the baghouse, conveyed, stored and hauled separately since its potential use is different.

- Feedwater treatment;
- Deaeration;
- Pumping; and
- Water or steam piping.

These items were not included because the AFBC boiler is being installed in addition to two existing gas- and oil-fired boilers. A booster feed pump and steam pressure reducing valve were included to accommodate the existing header pressures.

Operating costs include all raw materials, labor, utilities, consumable materials, repair, maintenance, and waste materials handling. They are based on the District of Columbia area. Unit costs and other considerations are listed below:

- Boiler efficiency - FBC = 82.5 percent with 4.1 percent carbon loss
  - Conventional = 84 percent with 2.2 percent carbon loss
- Coal - noncompliance high sulfur, \$40/ton (Eastern, 3.5 percent S, 8 percent ash, HHV = 12,500 Btu/lb)
  - compliance low sulfur, \$53/ton (Eastern, 0.7 percent S, 8 percent ash, HHV = 12,250 Btu/lb)
- Limestone (Ca/S = 3), \$15/ton
- Electricity, \$0.035/kwh
- Labor (average), \$8.00/man-hour
- Annual fixed charges = 18 percent of total capital cost (to include depreciation, interest, local taxes, and insurance).

The costs developed for the FBC burning high sulfur coal and the conventional system burning low sulfur compliance coal were considered in this analysis. Adjustments made to these costs to provide comparability are shown in Table B-2.

TABLE B-2. ADJUSTMENTS MADE TO A.G. McKEE COST BASIS

1. Total annual costs were developed based on use of the Eastern high sulfur coal noted for this study; i.e., 11,800 Btu/lb and \$17/ton. For FBC (82.5 percent efficiency) this converts to \$0.87/10^6 Btu output.

2. Eastern high sulfur coal was substituted for the compliance coal burned by the conventional boiler with ESP. This equates to a coal cost of \$0.86/10^6 Btu output based on 84 percent boiler efficiency.

3. A load factor of 0.6 was used to determine annualized capital costs.

These adjustments resulted in the following total annual costs:

- FBC boiler burning high sulfur coal - \$4.71/10^6 Btu output
- Conventional boiler burning high sulfur coal with ESP - \$4.34/10^6 Btu output

The ESP cost was not itemized, so that it was not subtracted from the conventional boiler cost.

APPENDIX C

DETAILED ENERGY AND COST TABULATIONS

The values presented in Tables C-6 through C-30 are calculated based on information from Appendix A; Tables C-1 through C-5, and from the PEDCo study of conventional boiler costs. Derivation of this background information is discussed in Chapters 3.0 and 4.0.

The background information is collated by computer to insure internal consistency under all options considered. The input to the program includes standard boiler costs, load factor, the coal analysis, drying requirements, and sulfur control information such as Ca/S and control level. This information is manipulated through mass and energy balances to determine input and output streams. These balances are then input to a costing subroutine to derive estimates of the effect on capital and operating cost for each boiler size.

The mass, energy and costing subroutines are the source of all final energy and cost estimates presented in Chapters 4.0 and 5.0. Additional information, such as SO_2 emitted, flue gas rates, and land use impact estimates, are printed out as needed for other chapters. Complete listings of all output are not included, for the sake of brevity. Sufficient information is included in Tables C-1 through C-30 to permit independent derivation of information presented.

TABLE C-1. COAL ANALYSES*

Coal	Moisture	Carbon	Hydrogen	Sulfur	Oxygen	Nitrogen	Ash	Btu/lb
Eastern high sulfur	8.79	64.80	4.43	3.50	6.56	1.30	10.58	11,800
Eastern low sulfur	2.83	78.75	4.71	0.90	4.91	1.50	6.90	13,800
Western low sulfur	20.80	57.60	3.20	0.60	11.20	1.20	5.40	9,600

*These values are averages developed from coals listed in Babcock & Wilcox "Useful Tables for Engineers and Steam Users," 12 ed., 1972.

TABLE C-2. PHYSICAL CONSTANTS

SENSIBLE HEAT		
Spent Residue	0.217	Btu/lb - °F
N_2	7	Btu/lb-mole-°F
O_2	7	Btu/lb-mole-°F
CO_2	9	Btu/lb-mole-°F
SO_2	9	Btu/lb-mole-°F
H_2O	8	Btu/lb-mole-°F
LATENT HEAT		
H_2O	1040	Btu/lb - °F
HEAT OF REACTION		
$CaCO_3 \longrightarrow CaO + CO_2$	1367	Btu/lb
$CaO + SO_2 + 1/2\ O_2 \longrightarrow CaSO_4$	3729	Btu/lb

TABLE C-3. BASE CONDITIONS AND RANGE FOR SENSITIVITY ANALYSIS

Parameter	Conventional combustion parametric considerations				FBC parametric considerations		
	Boiler capacity - MW_t						
	8.8	22	44	58.6	Std Condition	ITAR	Sensitivity
Excess air, %	50	50	50	30	20	20	0 - 100
Combustion efficiency, %	97	97	97	99	97	97	80 - 100
Ca/S ratio, m/m	-	-	-	-	3.5	0.6 - 3.5	1 - 10
SO_2 control efficiency, %	-	-	-	-	90	56 - 90	70 - 95
Coal sulfur, %	0.6 - 3.5	0.6 - 3.5	0.6 - 3.5	0.6 - 3.5	3.5	0.6 - 3.5	3.5
Coal HHV, Btu/lb	9,600 - 13,800	9,600 - 13,800	9,600 - 13,800	9,600 - 13,800	11,800	9,600 - 13,800	11,800
Coal cost, $/ton	6.75 - 29	6.75 - 29	6.75 - 29	6.75 - 29	17	6.75 - 29	8 - 60
Limestone cost, $/ton	-	-	-	-	8.00	8.00	5 - 30
Spent solids disposal, $/ton	-	-	-	-	40	40	5 - 40
Spent solids heat recovery, %	0	0	0	0	0	0	0 - 100
Spent solids temperature, °F	1,700	1,700	1,700	1,700	1,500	1,500	1,550 - 300
Flue gas temperature, °F	350 - 400	350 - 400	350 - 400	350 - 400	350	350	350
Ambient Air at FD fan, °F	80	80	80	80	80	80	80
Bottom ash, %	75	75	35	20	90	90	90
Load factor, %	60	60	60	60	60	60	30 - 100

TABLE C-4. INPUT PARAMETERS

HV, CAS, SC, ASH, DELT, C, H, S, O, AN, H2O

SHL1, SHL, CCOST, HLP, HPC

LOWW1, LOSS2, LOSS3, XS, XA, XSP, AV

HV - HEATING VALUE 11800 Btu/lb ⟶ 0.0118

CAS - CALCIUM TO SULFUR RATIO m/m

SC - SULFUR CONTROL, %

ASH - COAL ASH CONTENT, %/100

DELT - TEMPERATURE DIFFERENTIAL OF FLUE GAS, °F

C, H, S, O, AN - CARBON, HYDROGEN, SULFUR, OXYGEN, NITROGEN IN COAL, %/100

H2O - SURFACE MOISTURE REMOVAL REQUIREMENT, %/100

SHL1, SHL - TEMPERATURE DIFFERENTIAL FOR SOLIDS HEAT LOSS IN CONVENTIONAL

CCOST, HLP, HPC - COAL COST, LIMESTONE COST, DISPOSAL COST, $/TON

LOSS1, LOSS2, LOSS3 - CARBON LOSS FROM AFB, P.C., STOKERS, %

XSS, XSA, XSP - EXCESS AIR IN STOKERS, AFBC, P.C., %/100

AV - PLANT AVAILABILITY, %/100

TABLE C-5. POWER REQUIREMENTS FOR GAS MOVEMENT IN UNCONTROLLED AFBC AND CONVENTIONAL INDUSTRIAL BOILERS

Fan type	System components through which air (FD) or flue gas (ID) is conveyed		Typical pressure loss* - cm (in.) w.g.		Standard boiler capacity		Air and flue gas rates				Fan power requirements				Total FD and ID fan power requirements††			
	AFBC	CC	AFBC	CC	MW	(10^6 Btu/hr)	AFBC		CC		AFBC		CC		AFBC		CC	
							cms	acfm	cms	acfm	KW	(HP)	KW	(HP)	KW	HP	KW	HP
Forced draft	Air heater	Air heater	8.9 (3.5)	8.9 (3.5)	8.8	(30)	2.86	6,060	3.58	7,575	78.1	104.7	14.3	19.2				
	Ducts and steam coil heater	Ducts and steam coil heater	5.1 (2.0)	5.1 (2.0)	22	(75)	7.19	15,225	8.98	19,030	196	262.9	36.0	48.3				
	Plenum	Plenum	7.6 (3.0)	7.6 (3.0)	44	(150)	14.37	30,450	17.96	38,060	392.2	525.9	72.0	96.5				
	-	Burners	-	5.1 (2.0)**	58.6	(200)	19.33	40,950	20.94	44,365	527.4	707.2	138	185				
	Distribution plate		38.1 (15)†															
	Fluid bed		121.9 (48)†															
	Subtotal	Subtotal	181.6 (71.5)	26.7 (10.5)														
Induced draft	Freeboard	Furnace	0.3 (0.1)	0.3 (0.1)	8.8	(30)	4.72	10,000	5.90	12,500	23.9	32.1	19.4	26.0				
	-	Boiler and superheater	-	3.3 (1.3)	22	(75)	11.86	25,120	14.82	31,400	60.2	80.7	48.6	65.2				
	Primary cyclone	-	15.2 (6.0)‡	-	44	(150)	23.71	50,240	29.64	62,800	120	161	97	131				
	Economizer	Economizer	4.8 (1.9)	4.8 (1.9)	58.6	(200)	31.89	67,570	34.55	73,200	162	217	113	152				
	Air heater	Air heater	11.2 (4.4)	11.2 (4.4)														
	Flues	Flues	2.3 (0.9)	2.3 (0.9)														
	Subtotal	Subtotal	33.8 (13.3)	21.8 (8.6)														
	Total	Total			8.8	(30)									115	155	42	56
					22	(75)									287	386	91	122
					44	(150)									574	772	172	231
					58.6	(200)									766	1030	277	373

[*] Values for conventional boilers were taken from reference no. 6 in Section 5.

Estimated from reference no. 7 in Section 5. Other losses for AFBC were assumed equivalent to conventional boilers.

[‡] Estimated maximum draft loss through primary cyclone.

[§] Air and flue gas rates were derived from standard conventional boiler flue gas rates estimated by PEDCo. Volumetric rates are lower for AFBC to account for lower operating excess air ratio of 20 percent. Flue gas is at 177°C (350°F). Combustion air is at 27°C (80°F) for FD fan design.

[#] Fan power based on standard fan law, where: $HP = \frac{0.000157\ Q \times SP}{\text{fan efficiency}}$;

Q = acfm
SP = static pressure, in. w.g.
fan efficiency = 65 percent

$kw = HP \times 0.7457$

[**] Pressure loss of 15.2 cm (6 in.) was added for 58.6 MW_t PC boiler to account for primary air conveying coal to burners.

[††] Includes 10 percent contingency for ancillary air requirements.

TABLE C-6. POWER REQUIRED FOR LIMESTONE AND SPENT SOLIDS HANDLING, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	3.	16.	7.	40.	14.	80.	19.	107.
		LOW	4.2	3.	19.	7.	48.	14.	96.	19.	128.
		HIGH	2.3	3.	13.	7.	32.	14.	65.	19.	85.
	I 85%	AVERAGE	2.9	3.	15.	7.	37.	14.	75.	19.	97.
		LOW	3.8	3.	18.	7.	44.	14.	88.	19.	118.
		HIGH	2.1	3.	12.	7.	30.	14.	59.	19.	79.
	M 78.7%	AVERAGE	2.5	3.	13.	7.	33.	14.	66.	19.	87.
		LOW	3.4	3.	16.	7.	40.	14.	81.	19.	108.
		HIGH	1.8	3.	11.	7.	27.	14.	54.	19.	72.
	SIP 56%	AVERAGE	1.0	3.	7.	7.	19.	14.	37.	19.	50.
		LOW	1.2	3.	8.	7.	20.	14.	41.	19.	54.
		HIGH	0.8	3.	7.	7.	17.	14.	34.	19.	45.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	2.	4.	4.	10.	8.	20.	10.	27.
		LOW	3.7	2.	5.	4.	12.	8.	24.	10.	31.
		HIGH	2.0	2.	3.	4.	9.	8.	17.	10.	23.
	M 75%	AVERAGE	2.2	2.	4.	4.	9.	8.	18.	10.	24.
		LOW	3.2	2.	4.	4.	11.	8.	21.	10.	29.
		HIGH	1.6	2.	3.	4.	8.	8.	16.	10.	21.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	2.	4.	4.	10.	9.	20.	12.	27.
		LOW	3.6	2.	5.	4.	12.	9.	23.	12.	31.
		HIGH	2.0	2.	4.	4.	9.	9.	18.	12.	24.
	M 75%	AVERAGE	2.2	2.	4.	4.	9.	9.	18.	12.	24.
		LOW	3.2	2.	4.	4.	11.	9.	22.	12.	29.
		HIGH	1.6	2.	3.	4.	8.	9.	16.	12.	22.

TABLE C-7. POWER REQUIRED FOR COAL HANDLING, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	6.	6.	12.	12.	22.	22.	575.	29.
		LOW	4.2	6.	6.	12.	12.	22.	22.	575.	29.
		HIGH	2.3	6.	6.	12.	12.	22.	22.	575.	29.
	I 85%	AVERAGE	2.9	6.	6.	12.	12.	22.	22.	575.	29.
		LOW	3.8	6.	6.	12.	12.	22.	22.	575.	29
		HIGH	2.1	6.	6.	12.	12.	22.	22.	575.	29.
	M 78.7%	AVERAGE	2.5	6.	6.	12.	12.	22.	22.	575.	29.
		LOW	3.4	6.	6.	12.	12.	22.	22.	575.	29.
		HIGH	1.8	6.	6.	12.	12.	22.	22.	575.	29.
	SIP 56%	AVERAGE	1.0	6.	6.	12.	12.	22.	22.	575.	29.
		LOW	1.2	6.	6.	12.	12.	22.	22.	575.	29.
		HIGH	0.8	6.	6.	12.	12.	22.	22.	575.	29.
EASTERN LOW SULFUR (0.9% S)	S/I 85.9%	AVERAGE	2.8	5.	5.	11.	11.	19.	19.	25.	25.
		LOW	3.7	5.	5.	11.	11.	19.	19.	25.	25.
		HIGH	2.0	5.	5.	11.	11.	19.	19.	25.	25.
	M 75%	AVERAGE	2.2	5.	5.	11.	11.	19.	19.	25.	25.
		LOW	3.2	5.	5.	11.	11.	19.	19.	25.	25.
		HIGH	1.6	5.	5.	11.	11.	19.	19.	25.	25.
SUBBITUMINOUS LOW SULFUR (0.8% S)	S/I 85.2%	AVERAGE	2.7	7.	7.	14.	14.	27.	27.	1796.	55.
		LOW	3.6	7.	7.	14.	14.	27.	27.	1796.	55.
		HIGH	2.0	7.	7.	14.	14.	27.	27.	1796.	55.
	M 75%	AVERAGE	2.2	7.	7.	14.	14.	27.	27.	1796.	55.
		LOW	3.2	7.	7.	14.	14.	27.	27.	1796.	55.
		HIGH	1.6	7.	7.	14.	14.	27.	27.	1796.	55.

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TABLE C-8. POWER REQUIRED FOR BOILER FEEDWATER PUMPING, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	4.2	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	2.3	18.	18.	47.	47.	94.	94.	125.	125.
	I 85%	AVERAGE	2.9	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	3.8	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	2.1	18.	18.	47.	47.	94.	94.	125.	125.
	M 78.7%	AVERAGE	2.5	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	3.4	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	1.8	18.	18.	47.	47.	94.	94.	125.	125.
	SIP 56%	AVERAGE	1.0	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	1.2	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	0.8	18.	18.	47.	47.	94.	94.	125.	125.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	3.7	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	2.0	18.	18.	47.	47.	94.	94.	125.	125.
	M 75%	AVERAGE	2.2	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	3.2	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	1.6	18.	18.	47.	47.	94.	94.	125.	125.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	3.6	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	2.0	18.	18.	47.	47.	94.	94.	125.	125.
	M 75%	AVERAGE	2.2	18.	18.	47.	47.	94.	94.	125.	125.
		LOW	3.2	18.	18.	47.	47.	94.	94.	125.	125.
		HIGH	1.6	18.	18.	47.	47.	94.	94.	125.	125.

TABLE C-9. POWER REQUIRED FOR FANS, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.5	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	4.2	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	2.3	42.	115.	91.	287.	172.	574.	227.	766.
	I 85%	AVERAGE	2.9	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	3.8	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	2.1	42.	115.	91.	287.	172.	574.	227.	766.
	M 78.7%	AVERAGE	2.5	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	3.4	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	1.8	42.	115.	91.	287.	172.	574.	227.	766.
	SIP 56%	AVERAGE	1.0	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	1.2	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	0.8	42.	115.	91.	287.	172.	574.	227.	766.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	3.7	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	2.0	42.	115.	91.	287.	172.	574.	227.	766.
	M 75%	AVERAGE	2.2	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	3.2	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	1.6	42.	115.	91.	287.	172.	574.	227.	766.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	3.6	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	2.0	42.	115.	91.	287.	172.	574.	227.	766.
	M 75%	AVERAGE	2.2	42.	115.	91.	287.	172.	574.	227.	766.
		LOW	3.2	42.	115.	91.	287.	172.	574.	227.	766.
		HIGH	1.6	42.	115.	91.	287.	172.	574.	227.	766.

TABLE C-10. SOLIDS HEAT LOSS, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	24.	143.	61.	359.	72.	717.	72.	956.
		LOW	4.2	24.	213.	61.	533.	72.	1066.	72.	1421.
		HIGH	2.3	24.	66.	61.	165.	72.	350.	72.	459.
	I 85%	AVERAGE	2.9	24.	120.	61.	300.	72.	600.	72.	800.
		LOW	3.8	24.	190.	61.	474.	72.	949.	72.	1265.
		HIGH	2.1	24.	58.	61.	145.	72.	290.	72.	587.
	M 78.7%	AVERAGE	2.5	24.	99.	61.	246.	72.	493.	72.	657.
		LOW	3.4	24.	168.	61.	421.	72.	842.	72.	1122.
		HIGH	1.8	24.	44.	61.	111.	72.	222.	72.	295.
	SIP 56%	AVERAGE	1.0	24.	17.	61.	42.	72.	84.	72.	112.
		LOW	1.2	24.	32.	61.	81.	72.	161.	72.	215.
		HIGH	0.8	24.	1.	61.	3.	72.	6.	72.	8.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	13.	33.	33.	83.	39.	165.	37.	220.
		LOW	3.7	13.	48.	33.	120.	39.	241.	37.	321.
		HIGH	2.0	13.	20.	33.	49.	39.	98.	37.	150.
	M 75%	AVERAGE	2.2	13.	26.	33.	65.	39.	129.	37.	172.
		LOW	3.2	13.	43.	33.	107.	39.	213.	37.	285.
		HIGH	1.6	13.	16.	33.	39.	39.	79.	37.	105.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	15.	32.	38.	81.	43.	162.	42.	217.
		LOW	3.6	15.	47.	38.	118.	43.	235.	42.	313.
		HIGH	2.0	15.	21.	38.	53.	43.	106.	42.	141.
	M 75%	AVERAGE	2.2	15.	27.	38.	68.	43.	135.	42.	180.
		LOW	3.2	15.	43.	38.	108.	43.	216.	42.	288.
		HIGH	1.6	15.	17.	38.	43.	43.	87.	42.	115.

TABLE C-11. FLUE GAS HEAT LOSSES, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	1277.	955.	3192.	2388.	6384.	4777.	7381.	6369.
		LOW	4.2	1277.	962.	3192.	2404.	6384.	4809.	7381.	6412.
		HIGH	2.3	1277.	948.	3192.	2370.	6384.	4741.	7381.	6321.
	I 85%	AVERAGE	2.9	1277.	953.	3192.	2381.	6384.	4763.	7381.	6350.
		LOW	3.8	1277.	959.	3192.	2397.	6384.	4795.	7381.	6393.
		HIGH	2.1	1277.	947.	3192.	2367.	6384.	4734.	7381.	6312.
	M 78.7%	AVERAGE	2.5	1277.	950.	3192.	2375.	6384.	4749.	7381.	6332.
		LOW	3.4	1277.	956.	3192.	2391.	6384.	4781.	7381.	6375.
		HIGH	1.8	1277.	945.	3192.	2362.	6384.	4724.	7381.	6299.
	SIP 56%	AVERAGE	1.0	1277.	940.	3192.	2349.	6384.	4698.	7381.	6264.
		LOW	1.2	1277.	941.	3192.	2353.	6384.	4705.	7381.	6274.
		HIGH	0.8	1277.	938.	3192.	2345.	6384.	4691.	7381.	6255.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	1065.	883.	2664.	2207.	5327.	4415.	6317.	5886.
		LOW	3.7	1065.	884.	2664.	2211.	5327.	4422.	6317.	5896.
		HIGH	2.0	1065.	882.	2664.	2204.	5327.	4408.	6317.	5878.
	M 75%	AVERAGE	2.2	1065.	882.	2664.	2205.	5327.	4410.	6317.	5880.
		LOW	3.2	1065.	884.	2664.	2209.	5327.	4418.	6317.	5891.
		HIGH	1.6	1065.	881.	2664.	2203.	5327.	4405.	6317.	5874.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	1270.	1074.	3176.	2685.	6351.	5370.	6506.	7160.
		LOW	3.6	1270.	1075.	3176.	2688.	6351.	5377.	6506.	7169.
		HIGH	2.0	1270.	1073.	3176.	2682.	6351.	5364.	6506.	7153.
	M 75%	AVERAGE	2.2	1270.	1073.	3176.	2683.	6351.	5366.	6506.	7155.
		LOW	3.2	1270.	1075.	3176.	2687.	6351.	5374.	6506.	7165.
		HIGH	1.6	1270.	1072.	3176.	2681.	6351.	5362.	6506.	7149.

TABLE C-12. COMBUSTION LOSSES, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	4.2	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	2.3	264.	264.	659.	659.	1318.	1318.	586.	1757.
	I 85%	AVERAGE	2.9	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	3.8	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	2.1	264.	264.	659.	659.	1318.	1318.	586.	1757.
	M 78.7%	AVERAGE	2.5	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	3.4	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	1.8	264.	264.	659.	659.	1318.	1318.	586.	1757.
	SIP 56%	AVERAGE	1.0	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	1.2	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	0.8	264.	264.	659.	659.	1318.	1318.	586.	1757.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	3.7	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	2.0	264.	264.	659.	659.	1318.	1318.	586.	1757.
	M 75%	AVERAGE	2.2	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	3.2	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	1.6	264.	264.	659.	659.	1318.	1318.	586.	1757.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	3.6	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	2.0	264.	264.	659.	659.	1318.	1318.	586.	1757.
	M 75%	AVERAGE	2.2	264.	264.	659.	659.	1318.	1318.	586.	1757.
		LOW	3.2	264.	264.	659.	659.	1318.	1318.	586.	1757.
		HIGH	1.6	264.	264.	659.	659.	1318.	1318.	586.	1757.

TABLE C-13. RADIATIVE AND OTHER ENERGY LOSSES, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	4.2	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	2.3	265.	265.	479.	479.	750.	750.	905.	905.
	I 85%	AVERAGE	2.9	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	3.8	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	2.1	265.	265.	479.	479.	750.	750.	905.	905.
	M 78.7%	AVERAGE	2.5	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	3.4	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	1.8	265.	265.	479.	479.	750.	750.	905.	905.
	SIP 56%	AVERAGE	1.0	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	1.2	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	0.8	265.	265.	479.	479.	750.	750.	905.	905.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	3.7	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	2.0	265.	265.	479.	479.	750.	750.	905.	905.
	M 75%	AVERAGE	2.2	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	3.2	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	1.6	265.	265.	479.	479.	750.	750.	905.	905.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	3.6	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	2.0	265.	265.	479.	479.	750.	750.	905.	905.
	M 75%	AVERAGE	2.2	265.	265.	479.	479.	750.	750.	905.	905.
		LOW	3.2	265.	265.	479.	479.	750.	750.	905.	905.
		HIGH	1.6	265.	265.	479.	479.	750.	750.	905.	905.

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TABLE C-14. AUXILIARY POWER REQUIREMENTS, kW

				BOILER CAPACITY-MW							
				8.8		22		44		58.6	
COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	70.	155.	157.	386.	302.	771.	399.	1027.
		LOW	4.2	70.	158.	157.	394.	302.	786.	399.	1047.
		HIGH	2.3	70.	152.	157.	378.	302.	754.	399.	1004.
	I 85%	AVERAGE	2.9	70.	154.	157.	382.	302.	765.	399.	1017.
		LOW	3.8	70.	157.	157.	390.	302.	776.	399.	1037.
		HIGH	2.1	70.	151.	157.	376.	302.	750.	399.	999.
	M 78.7%	AVERAGE	2.5	70.	152.	157.	379.	302.	756.	399.	1007.
		LOW	3.4	70.	155.	157.	386.	302.	771.	399.	1027.
		HIGH	1.8	70.	150.	157.	373.	302.	744.	399.	991.
	SIP 56%	AVERAGE	1.0	70.	147.	157.	365.	302.	728.	399.	970.
		LOW	1.2	70.	147.	157.	366.	302.	731.	399.	974.
		HIGH	0.8	70.	146.	157.	363.	302.	724.	399.	965.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	68.	143.	152.	354.	293.	707.	387.	943.
		LOW	3.7	68.	143.	152.	356.	293.	711.	387.	947.
		HIGH	2.0	68.	142.	152.	353.	293.	705.	387.	939.
	M 75%	AVERAGE	2.2	68.	142.	152.	353.	293.	705.	387.	940.
		LOW	3.2	68.	143.	152.	355.	293.	709.	387.	944.
		HIGH	1.6	68.	142.	152.	352.	293.	703.	387.	937.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	69.	144.	157.	358.	302.	715.	399.	953.
		LOW	3.6	69.	145.	157.	360.	302.	718.	399.	957.
		HIGH	2.0	69.	144.	157.	357.	302.	713.	399.	950.
	M 75%	AVERAGE	2.2	69.	144.	157.	357.	302.	713.	399.	950.
		LOW	3.2	69.	144.	157.	359.	302.	717.	399.	955.
		HIGH	1.6	69.	143.	157.	356.	302.	711.	399.	948.

TABLE C-15. TOTAL INHERENT ENERGY LOSSES, kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (5.5% S)	S 90%	AVERAGE	3.3	1830.	1628.	4391.	3885.	8524.	7562.	9286.	9986.
		LOW	4.2	1830.	1704.	4391.	4076.	8524.	7943.	9286.	10493.
		HIGH	2.5	1830.	1543.	4391.	3674.	8524.	7139.	9286.	9421.
	I 85%	AVERAGE	2.9	1830.	1601.	4391.	3820.	8524.	7431.	9286.	9811.
		LOW	3.8	1830.	1678.	4391.	4010.	8524.	7812.	9286.	10319.
		HIGH	2.1	1830.	1534.	4391.	3651.	8524.	7092.	9286.	9360.
	M 78.7%	AVERAGE	2.5	1830.	1577.	4391.	3759.	8524.	7310.	9286.	9650.
		LOW	3.4	1830.	1653.	4391.	3950.	8524.	7691.	9286.	10158.
		HIGH	1.8	1830.	1518.	4391.	3611.	8524.	7014.	9286.	9255.
	SIP 56%	AVERAGE	1.0	1830.	1485.	4391.	3529.	8524.	6650.	9286.	9036.
		LOW	1.2	1830.	1502.	4391.	3572.	8524.	6934.	9286.	9149.
		HIGH	0.8	1830.	1468.	4391.	3487.	8524.	6765.	9286.	8923.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	1608.	1445.	3835.	3428.	7434.	6648.	7843.	8767.
		LOW	3.7	1608.	1461.	3835.	3470.	7434.	6731.	7843.	8878.
		HIGH	2.0	1608.	1430.	3835.	3391.	7434.	6574.	7843.	8668.
	M 75%	AVERAGE	2.2	1608.	1437.	3835.	3408.	7434.	6607.	7843.	8713.
		LOW	3.2	1608.	1455.	3835.	3454.	7434.	6700.	7843.	8836.
		HIGH	1.6	1608.	1426.	3835.	3380.	7434.	6552.	7843.	8639.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	1814.	1635.	4351.	3904.	8463.	7600.	9798.	10057.
		LOW	3.6	1814.	1651.	4351.	3944.	8463.	7680.	9798.	10145.
		HIGH	2.0	1814.	1625.	4351.	3874.	8463.	7536.	9798.	9954.
	M 75%	AVERAGE	2.2	1814.	1629.	4351.	3889.	8463.	7569.	9798.	9995.
		LOW	3.2	1814.	1647.	4351.	3933.	8463.	7658.	9798.	10113.
		HIGH	1.6	1814.	1618.	4351.	3862.	8463.	7516.	9798.	9925.

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TABLE C-16. TOTAL ENERGY LOSSES (AUXILIARY PLUS INHERENT), kW

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	1899.	1783.	4548.	4271.	8826.	8332.	9685.	11012.
		LOW	4.2	1899.	1862.	4548.	4469.	8826.	8728.	9685.	11541.
		HIGH	2.3	1899.	1695.	4548.	4051.	8826.	7892.	9685.	10426.
	I 85%	AVERAGE	2.9	1899.	1755.	4548.	4202.	8826.	8194.	9685.	10828.
		LOW	3.8	1899.	1834.	4548.	4400.	8826.	8590.	9685.	11356.
		HIGH	2.1	1899.	1685.	4548.	4026.	8826.	7842.	9685.	10359.
	M 76.7%	AVERAGE	2.5	1899.	1730.	4548.	4138.	8826.	8066.	9685.	10657.
		LOW	3.4	1899.	1809.	4548.	4336.	8826.	8462.	9685.	11185.
		HIGH	1.8	1899.	1668.	4548.	3984.	8826.	7758.	9685.	10246.
	SIP 56%	AVERAGE	1.0	1899.	1632.	4548.	3894.	8826.	7577.	9685.	10006.
		LOW	1.2	1899.	1649.	4548.	3938.	8826.	7665.	9685.	10123.
		HIGH	0.8	1899.	1614.	4548.	3850.	8826.	7489.	9685.	9889.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	1675.	1587.	3988.	3783.	7727.	7355.	8230.	9710.
		LOW	3.7	1675.	1605.	3988.	3826.	7727.	7442.	8230.	9825.
		HIGH	2.0	1675.	1572.	3988.	3744.	7727.	7278.	8230.	9607.
	M 75%	AVERAGE	2.2	1675.	1579.	3988.	3761.	7727.	7512.	8230.	9652.
		LOW	3.2	1675.	1598.	3988.	3809.	7727.	7408.	8230.	9780.
		HIGH	1.6	1675.	1567.	3988.	3732.	7727.	7255.	8230.	9576.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	1884.	1779.	4508.	4263.	8764.	8315.	10196.	10990.
		LOW	3.6	1884.	1796.	4508.	4304.	8764.	8398.	10196.	11100.
		HIGH	2.0	1884.	1767.	4508.	4231.	8764.	8251.	10196.	10904.
	M 75%	AVERAGE	2.2	1884.	1773.	4508.	4246.	8764.	8282.	10196.	10946.
		LOW	3.2	1884.	1791.	4508.	4292.	8764.	8374.	10196.	11068.
		HIGH	1.6	1884.	1762.	4508.	4219.	8764.	8227.	10196.	10872.

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TABLE C-17. STATION EFFICIENCY, PERCENT

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW 8.6 CONVENTIONAL	8.6 AFBC	22 CONVENTIONAL	22 AFBC	44 CONVENTIONAL	44 AFBC	58.6 CONVENTIONAL	58.6 AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	78.38	79.71	79.30	80.56	79.91	81.63	83.47	81.20
		LOW	4.2	78.38	78.81	79.30	79.65	79.91	80.13	83.47	80.30
		HIGH	2.5	78.38	80.71	79.30	81.56	79.91	82.03	83.47	82.20
	I 85%	AVERAGE	2.9	78.38	80.02	79.30	80.87	79.91	81.35	83.47	81.51
		LOW	3.8	78.38	79.12	79.30	79.97	79.91	80.45	83.47	80.61
		HIGH	2.1	78.38	80.82	79.30	81.67	79.91	82.15	83.47	82.32
	M 78.7%	AVERAGE	2.5	78.38	80.52	79.30	81.16	79.91	81.64	83.47	81.81
		LOW	3.4	78.38	79.41	79.30	80.26	79.91	80.74	83.47	80.90
		HIGH	1.8	78.38	81.02	79.30	81.86	79.91	82.34	83.47	82.51
	SIP 56%	AVERAGE	1.0	78.38	81.43	79.30	82.27	79.91	82.75	83.47	82.92
		LOW	1.2	78.38	81.23	79.30	82.07	79.91	82.55	83.47	82.72
		HIGH	0.8	78.38	81.63	79.30	82.47	79.91	82.95	83.47	83.12
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	80.93	81.93	81.85	82.78	82.41	83.26	85.95	83.42
		LOW	3.7	80.93	81.74	81.85	82.58	82.41	83.06	85.95	83.23
		HIGH	2.0	80.93	82.11	81.85	82.95	82.41	83.43	85.95	83.60
	M 75%	AVERAGE	2.2	80.93	82.03	81.85	82.88	82.41	83.36	85.95	83.52
		LOW	3.2	80.93	81.81	81.85	82.66	82.41	83.14	85.95	83.30
		HIGH	1.6	80.93	82.16	81.85	83.01	82.41	83.49	85.95	83.65
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	78.56	79.75	79.48	80.59	80.05	81.07	82.59	81.24
		LOW	3.6	78.56	79.56	79.48	80.41	80.05	80.86	82.59	81.05
		HIGH	2.0	78.56	79.89	79.48	80.74	80.05	81.22	82.59	81.38
	M 75%	AVERAGE	2.2	78.56	79.82	79.48	80.67	80.05	81.15	82.59	81.31
		LOW	3.2	78.56	79.61	79.48	80.46	80.05	80.94	82.59	81.10
		HIGH	1.6	78.56	79.95	79.48	80.79	80.05	81.27	82.59	81.44

TABLE C-18. KW/KG SO_2 REMOVED

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW			
				8.8	22	44	58.6
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	-3.16	-3.00	-2.68	5.41
		LOW	4.2	-1.01	-0.85	-0.53	7.56
		HIGH	2.3	-5.56	-5.39	-5.07	3.02
	I 85%	AVERAGE	2.9	-4.15	-3.97	-3.64	4.95
		LOW	3.8	-1.87	-1.70	-1.36	7.21
		HIGH	2.1	-6.17	-6.00	-5.66	2.91
	M 78.7%	AVERAGE	2.5	-5.27	-5.09	-4.73	4.53
		LOW	3.4	-2.81	-2.63	-2.26	6.99
		HIGH	1.8	-7.19	-7.00	-6.64	2.61
	SIP 56%	AVERAGE	1.0	-11.68	-11.42	-10.90	2.10
		LOW	1.2	-10.91	-10.65	-10.14	2.87
		HIGH	0.8	-12.44	-12.19	-11.67	1.33
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	-11.79	-10.99	-9.97	29.74
		LOW	3.7	-9.47	-8.68	-7.65	32.05
		HIGH	2.0	-13.84	-13.05	-12.03	27.68
	M 75%	AVERAGE	2.2	-14.47	-13.58	-12.44	31.98
		LOW	3.2	-11.59	-10.70	-9.56	34.86
		HIGH	1.6	-16.20	-15.31	-14.17	30.25
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	-14.68	-13.84	-12.67	16.79
		LOW	3.6	-12.34	-11.50	-10.33	19.12
		HIGH	2.0	-16.49	-15.65	-14.46	14.97
	M 75%	AVERAGE	2.2	-17.32	-16.39	-15.06	17.59
		LOW	3.2	-14.44	-13.51	-12.21	20.46
		HIGH	1.6	-19.04	-18.11	-16.81	15.86

TABLE C-19. BOILER EFFICIENCY PERCENT

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	79.17	81.48	80.01	82.31	80.60	82.79	84.15	82.95
		LOW	4.2	79.17	80.61	80.01	81.45	80.60	81.92	84.15	82.09
		HIGH	2.3	79.17	82.44	80.01	83.28	80.60	83.75	84.15	83.92
	I 85%	AVERAGE	2.9	79.17	81.77	80.01	82.61	80.60	83.09	84.15	83.25
		LOW	3.8	79.17	80.91	80.01	81.74	80.60	82.22	84.15	82.38
		HIGH	2.1	79.17	82.54	80.01	83.38	80.60	83.86	84.15	84.02
	M 78.7%	AVERAGE	2.5	79.17	82.05	80.01	82.89	80.60	83.36	84.15	83.53
		LOW	3.4	79.17	81.18	80.01	82.02	80.60	82.49	84.15	82.66
		HIGH	1.8	79.17	82.72	80.01	83.56	80.60	84.03	84.15	84.20
	SIP 56%	AVERAGE	1.0	79.17	83.10	80.01	83.93	80.60	84.41	84.15	84.57
		LOW	1.2	79.17	82.90	80.01	83.74	80.60	84.22	84.15	84.38
		HIGH	0.8	79.17	83.29	80.01	84.13	80.60	84.60	84.15	84.77
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	81.70	83.56	82.54	84.39	83.08	84.87	86.61	85.03
		LOW	3.7	81.70	83.37	82.54	84.20	83.08	84.68	86.61	84.84
		HIGH	2.0	81.70	83.72	82.54	84.56	83.08	85.04	86.61	85.20
	M 75%	AVERAGE	2.2	81.70	83.65	82.54	84.49	83.08	84.96	86.61	85.13
		LOW	3.2	81.70	83.44	82.54	84.27	83.08	84.75	86.61	84.92
		HIGH	1.6	81.70	83.77	82.54	84.61	83.08	85.09	86.61	85.25
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	79.35	81.39	80.19	82.22	80.74	82.70	83.27	82.86
		LOW	3.6	79.35	81.21	80.19	82.04	80.74	82.52	83.27	82.66
		HIGH	2.0	79.35	81.53	80.19	82.37	80.74	82.84	83.27	83.01
	M 75%	AVERAGE	2.2	79.35	81.46	80.19	82.30	80.74	82.77	83.27	82.94
		LOW	3.2	79.35	81.26	80.19	82.09	80.74	82.57	83.27	82.73
		HIGH	1.6	79.35	81.58	80.19	82.42	80.74	82.89	83.27	83.06

TABLE C-20. TOTAL TURNKEY COST FOR LIMESTONE STORAGE AND HANDLING - DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	0.	88453.	0.	192208.	0.	523405.	0.	697874.
		LOW	4.2	0.	109899.	0.	333076.	0.	666152.	0.	888203.
		HIGH	2.3	0.	63278.	0.	144145.	0.	364798.	0.	486597.
	I 85%	AVERAGE	2.9	0.	78553.	0.	174045.	0.	459962.	0.	613283.
		LOW	3.8	0.	100509.	0.	301355.	0.	602709.	0.	803613.
		HIGH	2.1	0.	58073.	0.	133469.	0.	333076.	0.	444102.
	M 78.7%	AVERAGE	2.5	0.	68427.	0.	154465.	0.	346519.	0.	528692.
		LOW	3.4	0.	90893.	0.	196527.	0.	534266.	0.	719022.
		HIGH	1.8	0.	50160.	0.	116793.	0.	285494.	0.	360659.
	SIP 56%	AVERAGE	1.0	0.	28433.	0.	68426.	0.	127944.	0.	162796.
		LOW	1.2	0.	33950.	0.	81049.	0.	149309.	0.	187800.
		HIGH	0.8	0.	22860.	0.	55449.	0.	105232.	0.	135273.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	0.	17476.	0.	42705.	0.	82130.	0.	106591.
		LOW	3.7	0.	22982.	0.	55735.	0.	105743.	0.	135848.
		HIGH	2.0	0.	12536.	0.	30839.	0.	60003.	0.	78517.
	M 75%	AVERAGE	2.2	0.	13775.	0.	33830.	0.	65636.	0.	85714.
		LOW	3.2	0.	19929.	0.	48538.	0.	92792.	0.	119914.
		HIGH	1.6	0.	10050.	0.	24805.	0.	48538.	0.	65766.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	0.	16168.	0.	39579.	0.	76357.	0.	99320.
		LOW	3.6	0.	21458.	0.	52150.	0.	99320.	0.	127999.
		HIGH	2.0	0.	12019.	0.	24587.	0.	57637.	0.	75485.
	M 75%	AVERAGE	2.2	0.	13208.	0.	32461.	0.	63062.	0.	82430.
		LOW	3.2	0.	19113.	0.	46601.	0.	89268.	0.	115526.
		HIGH	1.6	0.	9635.	0.	23793.	0.	46601.	0.	61261.

TABLE C-21. TOTAL TURNKEY COST FOR SPENT SOLIDS STORAGE AND HANDLING - DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	18544.	86224.	45213.	187138.	42317.	509453.	32544.	679271.
		LOW	4.2	18544.	98425.	45213.	207982.	42317.	589713.	32544.	786284.
		HIGH	2.3	18544.	72225.	45213.	161220.	42317.	420276.	32544.	560367.
	I 85%	AVERAGE	2.9	18544.	79877.	45213.	175641.	42317.	468654.	32544.	624871.
		LOW	3.8	18544.	92270.	45213.	197680.	42317.	548915.	32544.	741884.
		HIGH	2.1	18544.	68545.	45213.	154077.	42317.	397311.	32544.	524749.
	M 78.7%	AVERAGE	2.5	18544.	73220.	45213.	163130.	42317.	426520.	32544.	568694.
		LOW	3.4	18544.	85811.	45213.	186404.	42317.	506760.	32544.	675707.
		HIGH	1.8	18544.	63168.	45213.	143403.	42317.	364096.	32544.	485461.
	SIP 56%	AVERAGE	1.0	18544.	47496.	45213.	110787.	42317.	195069.	32544.	359294.
		LOW	1.2	18544.	50490.	45213.	117185.	42317.	204240.	32544.	383075.
		HIGH	0.8	18544.	44483.	45213.	104274.	42317.	165435.	32544.	335513.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	10397.	25446.	25638.	61427.	23965.	115559.	18353.	147594.
		LOW	3.7	10397.	28506.	25638.	68502.	23965.	127798.	18353.	162213.
		HIGH	2.0	10397.	22712.	25638.	55044.	23965.	104306.	18353.	133935.
	M 75%	AVERAGE	2.2	10397.	23046.	25638.	55828.	23965.	105698.	18353.	135855.
		LOW	3.2	10397.	26460.	25638.	63779.	23965.	114655.	18353.	152516.
		HIGH	1.6	10397.	20987.	25638.	50991.	23965.	97061.	18353.	125039.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	11684.	26061.	28759.	62855.	26888.	118049.	20605.	150569.
		LOW	3.6	11684.	28990.	28759.	69616.	26888.	129701.	20605.	164463.
		HIGH	2.0	11684.	23772.	28759.	57525.	26888.	108703.	20605.	139296.
	M 75%	AVERAGE	2.2	11684.	24118.	28759.	58334.	26888.	110130.	20605.	141029.
		LOW	3.2	11684.	27383.	28759.	65915.	26888.	123352.	20605.	156932.
		HIGH	1.6	11684.	22149.	28759.	53724.	26888.	101955.	20605.	131057.

TABLE C-22. ANNUAL COST OF LIMESTONE PURCHASE - DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	0.	21681.	0.	54202.	0.	108405.	0.	144540.
		LOW	4.2	0.	27594.	0.	68985.	0.	137970.	0.	183960.
		HIGH	2.3	0.	15111.	0.	37777.	0.	75555.	0.	100740.
	I 85%	AVERAGE	2.9	0.	19053.	0.	47632.	0.	95265.	0.	127020.
		LOW	3.8	0.	24966.	0.	62415.	0.	124830.	0.	166440.
		HIGH	2.1	0.	13797.	0.	34492.	0.	68985.	0.	91980.
	4 78.7%	AVERAGE	2.5	0.	16425.	0.	41062.	0.	82125.	0.	109500.
		LOW	3.4	0.	22338.	0.	55845.	0.	111690.	0.	148920.
		HIGH	1.8	0.	11826.	0.	29565.	0.	59130.	0.	78840.
	SIP 56%	AVERAGE	1.0	0.	6570.	0.	16425.	0.	32850.	0.	43800.
		LOW	1.2	0.	7884.	0.	19710.	0.	39420.	0.	52560.
		HIGH	0.8	0.	5256.	0.	13140.	0.	26280.	0.	35040.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	0.	3999.	0.	9998.	0.	19996.	0.	26661.
		LOW	3.7	0.	5285.	0.	13211.	0.	26423.	0.	35230.
		HIGH	2.0	0.	2857.	0.	7141.	0.	14283.	0.	19043.
	M 75%	AVERAGE	2.2	0.	3142.	0.	7855.	0.	15711.	0.	20948.
		LOW	3.2	0.	4570.	0.	11426.	0.	22852.	0.	30470.
		HIGH	1.6	0.	2285.	0.	5713.	0.	11426.	0.	15235.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	0.	3696.	0.	9239.	0.	18478.	0.	24637.
		LOW	3.6	0.	4927.	0.	12319.	0.	24637.	0.	32850.
		HIGH	2.0	0.	2737.	0.	6844.	0.	13687.	0.	18250.
	M 75%	AVERAGE	2.2	0.	3011.	0.	7528.	0.	15056.	0.	20075.
		LOW	3.2	0.	4380.	0.	10950.	0.	21900.	0.	29200.
		HIGH	1.6	0.	2190.	0.	5475.	0.	10950.	0.	14600.

TABLE C-23. ANNUAL COST OF SPENT SOLIDS DISPOSAL - DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	21247.	105728.	53117.	264320.	49576.	528640.	37772.	704854.
		LOW	4.2	21247.	122384.	53117.	305961.	49576.	611925.	37772.	815847.
		HIGH	2.3	21247.	87221.	53117.	218052.	49576.	436104.	37772.	581472.
	I 85%	AVERAGE	2.9	21247.	97261.	53117.	243152.	49576.	486304.	37772.	648406.
		LOW	3.8	21247.	113917.	53117.	284793.	49576.	569587.	37772.	759449.
		HIGH	2.1	21247.	82455.	53117.	206138.	49576.	412275.	37772.	549701.
	M 78.7%	AVERAGE	2.5	21247.	88517.	53117.	221292.	49576.	442584.	37772.	590113.
		LOW	3.4	21247.	105173.	53117.	262934.	49576.	525867.	37772.	701156.
		HIGH	1.8	21247.	75562.	53117.	188904.	49576.	377809.	37772.	503745.
	SIP 56%	AVERAGE	1.0	21247.	55924.	53117.	139810.	49576.	279620.	37772.	372826.
		LOW	1.2	21247.	59625.	53117.	149063.	49576.	298127.	37772.	397502.
		HIGH	0.8	21247.	52222.	53117.	130556.	49576.	261112.	37772.	348150.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	11826.	29339.	29565.	73347.	27594.	146695.	21024.	195593.
		LOW	3.7	11826.	32960.	29565.	82400.	27594.	164800.	21024.	219733.
		HIGH	2.0	11826.	26120.	29565.	65301.	27594.	130602.	21024.	174136.
	M 75%	AVERAGE	2.2	11826.	26513.	29565.	66283.	27594.	132566.	21024.	176754.
		LOW	3.2	11826.	30536.	29565.	76341.	27594.	152682.	21024.	203577.
		HIGH	1.6	11826.	24099.	29565.	60248.	27594.	120496.	21024.	160661.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	13304.	30065.	33261.	75163.	31045.	150326.	23652.	200434.
		LOW	3.6	13304.	33535.	33261.	83838.	31045.	167676.	23652.	223568.
		HIGH	2.0	13304.	27366.	33261.	68415.	31045.	136831.	23652.	182441.
	M 75%	AVERAGE	2.2	13304.	27774.	33261.	69434.	31045.	138868.	23652.	185158.
		LOW	3.2	13304.	31629.	33261.	79073.	31045.	158147.	23652.	210862.
		HIGH	1.6	13304.	25460.	33261.	63651.	31045.	127301.	23652.	169735.

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TABLE C-24. ANNUAL COST OF ELECTRICITY - DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	9434.	21056.	21274.	52343.	41008.	104489.	54169.	139252.
		LOW	4.2	9434.	21470.	21274.	53378.	41008.	106558.	54169.	142011.
		HIGH	2.5	9434.	20596.	21274.	51194.	41008.	102189.	54169.	136167.
	I 85%	AVERAGE	2.9	9434.	20856.	21274.	51844.	41008.	103490.	54169.	137920.
		LOW	3.8	9434.	21270.	21274.	52876.	41008.	105559.	54169.	140679.
		HIGH	2.1	9434.	20489.	21274.	50924.	41008.	101651.	54169.	135468.
	M 78.7%	AVERAGE	2.5	9434.	20655.	21274.	51534.	41008.	102471.	54169.	136562.
		LOW	3.4	9434.	21066.	21274.	52369.	41008.	104540.	54169.	139320.
		HIGH	1.8	9434.	20331.	21274.	50530.	41008.	100861.	54169.	134416.
	SIP 56%	AVERAGE	1.0	9434.	19891.	21274.	49431.	41008.	98663.	54169.	131485.
		LOW	1.2	9434.	19983.	21274.	49660.	41008.	99123.	54169.	132098.
		HIGH	0.8	9434.	19799.	21274.	49201.	41008.	98203.	54169.	130872.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	9187.	19345.	20656.	48064.	39772.	95931.	52522.	127842.
		LOW	3.7	9187.	19434.	20656.	48289.	39772.	96381.	52522.	128441.
		HIGH	2.0	9187.	19265.	20656.	47864.	39772.	95531.	52522.	127308.
	M 75%	AVERAGE	2.2	9187.	19278.	20656.	47899.	39772.	95600.	52522.	127401.
		LOW	3.2	9187.	19378.	20656.	48149.	39772.	96100.	52522.	128067.
		HIGH	1.6	9187.	19218.	20656.	47749.	39772.	95300.	52522.	127001.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	9418.	19551.	21234.	48581.	40927.	96965.	54061.	129220.
		LOW	3.6	9418.	19638.	21234.	48797.	40927.	97396.	54061.	129795.
		HIGH	2.0	9418.	19484.	21234.	48414.	40927.	96629.	54061.	128773.
	M 75%	AVERAGE	2.2	9418.	19498.	21234.	48448.	40927.	96698.	54061.	128865.
		LOW	3.2	9418.	19594.	21234.	48687.	40927.	97177.	54061.	129503.
		HIGH	1.6	9418.	19441.	21234.	48304.	40927.	96411.	54061.	128482.

TABLE C-25. ANNUAL COAL PURCHASE COST - DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		LOW	4.2	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		HIGH	2.3	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
	I 85%	AVERAGE	2.9	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		LOW	3.8	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		HIGH	2.1	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
	M 78.7%	AVERAGE	2.5	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		LOW	3.4	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		HIGH	1.8	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
	SIP 56%	AVERAGE	1.0	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		LOW	1.2	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
		HIGH	0.8	113583.	113583.	283957.	283957.	567915.	567915.	757220.	757220.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	165678.	165678.	414195.	414195.	828391.	828391.	1104520.	1104520.
		LOW	3.7	165678.	165678.	414195.	414195.	828391.	828391.	1104520.	1104520.
		HIGH	2.0	165678.	165678.	414195.	414195.	828391.	828391.	1104520.	1104520.
	M 75%	AVERAGE	2.2	165678.	165678.	414195.	414195.	828391.	828391.	1104520.	1104520.
		LOW	3.2	165678.	165678.	414195.	414195.	828391.	828391.	1104520.	1104520.
		HIGH	1.6	165678.	165678.	414195.	414195.	828391.	828391.	1104520.	1104520.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	55434.	55434.	138586.	138586.	277172.	277172.	369563.	369563.
		LOW	3.6	55434.	55434.	138586.	138586.	277172.	277172.	369563.	369563.
		HIGH	2.0	55434.	55434.	138586.	138586.	277172.	277172.	369563.	369563.
	M 75%	AVERAGE	2.2	55434.	55434.	138586.	138586.	277172.	277172.	369563.	369563.
		LOW	3.2	55434.	55434.	138586.	138586.	277172.	277172.	369563.	369563.
		HIGH	1.6	55434.	55434.	138586.	138586.	277172.	277172.	369563.	369563.

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TABLE C-26. TOTAL ANNUAL COST, AFBC WITH SO_2 CONTROL AND UNCONTROLLED CONVENTIONAL BOILERS — DOLLARS

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	927071.	995651.	1826049.	2259707.	3044177.	3856745.	4035712.	4960527.
		LOW	4.2	927071.	1023547.	1826049.	2340775.	3044177.	4004221.	4035712.	5157161.
		HIGH	2.3	927071.	964394.	1826049.	2185063.	3044177.	3692864.	4035712.	4742046.
	I 85%	AVERAGE	2.9	927071.	981984.	1826049.	2227139.	3044177.	3785052.	4035712.	4864935.
		LOW	3.8	927071.	1009982.	1826049.	2306403.	3044177.	3932527.	4035712.	5061569.
		HIGH	2.1	927071.	956910.	1826049.	2166993.	3044177.	3653962.	4035712.	4690150.
	M 78.7%	AVERAGE	2.5	927071.	967958.	1826049.	2193515.	3044177.	3711758.	4035712.	4767212.
		LOW	3.4	927071.	996059.	1826049.	2260512.	3044177.	3859234.	4035712.	4963844.
		HIGH	1.8	927071.	945947.	1826049.	2140444.	3044177.	3597055.	4035712.	4614274.
	SIP 56%	AVERAGE	1.0	927071.	915153.	1826049.	2065287.	3044177.	3422715.	4035712.	4395155.
		LOW	1.2	927071.	921503.	1826049.	2080833.	3044177.	3452706.	4035712.	4436527.
		HIGH	0.8	927071.	908792.	1826049.	2049673.	3044177.	3392448.	4035712.	4353616.
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	921109.	905827.	1836259.	2067308.	3090555.	3437674.	4146445.	4413526.
		LOW	3.7	921109.	912074.	1836259.	2082734.	3090555.	3467690.	4146445.	4453049.
		HIGH	2.0	921109.	900266.	1836259.	2053540.	3090555.	3410594.	4146445.	4377625.
	M 75%	AVERAGE	2.2	921109.	901188.	1836259.	2055822.	3090555.	3415082.	4146445.	4383540.
		LOW	3.2	921109.	908136.	1836259.	2073009.	3090555.	3448842.	4146445.	4428008.
		HIGH	1.6	921109.	897012.	1836259.	2045471.	3090555.	3394670.	4146445.	4350582.
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 85.2%	AVERAGE	2.7	933862.	864712.	1763479.	1905824.	3028267.	3095923.	3999383.	3929040.
		LOW	3.6	933862.	870700.	1763479.	1920617.	3028267.	3124918.	3999383.	3967575.
		HIGH	2.0	933862.	860048.	1763479.	1894277.	3028267.	3073204.	3999383.	3894084.
	M 75%	AVERAGE	2.2	933862.	860967.	1763479.	1896552.	3028267.	3077680.	3999383.	3905384.
		LOW	3.2	933862.	867626.	1763479.	1913024.	3028267.	3110038.	3999383.	3948000.
		HIGH	1.6	933862.	856966.	1763479.	1886633.	3028267.	3058122.	3999383.	3874557.

TABLE C-27. TOTAL ANNUAL COST OF AFBC AND UNCONTROLLED CONVENTIONAL BOILERS, $/10^6 Btu OUTPUT

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	7.39	7.75	5.76	6.96	4.77	5.91	4.56	5.69
		LOW	4.2	7.39	8.04	5.76	7.28	4.77	6.19	4.56	5.97
		HIGH	2.3	7.39	7.42	5.76	6.66	4.77	5.60	4.56	5.38
	I 85%	AVERAGE	2.9	7.39	7.62	5.76	6.84	4.77	5.78	4.56	5.56
		LOW	3.8	7.39	7.91	5.76	7.15	4.77	6.06	4.56	5.84
		HIGH	2.1	7.39	7.36	5.76	6.60	4.77	5.53	4.56	5.31
	M 78.7%	AVERAGE	2.5	7.39	7.48	5.76	6.72	4.77	5.65	4.56	5.43
		LOW	3.4	7.39	7.78	5.76	6.99	4.77	5.93	4.56	5.71
		HIGH	1.8	7.39	7.26	5.76	6.51	4.77	5.44	4.56	5.22
	SIP 56%	AVERAGE	1.0	7.39	7.00	5.76	6.25	4.77	5.15	4.56	4.95
		LOW	1.2	7.39	7.06	5.76	6.31	4.77	5.21	4.56	5.01
		HIGH	0.8	7.39	6.93	5.76	6.19	4.77	5.10	4.56	4.90
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	7.12	6.87	5.62	6.21	4.70	5.13	4.55	4.93
		LOW	3.7	7.12	6.93	5.62	6.27	4.70	5.19	4.55	4.99
		HIGH	2.0	7.12	6.82	5.62	6.16	4.70	5.08	4.55	4.89
	M 75%	AVERAGE	2.2	7.12	6.83	5.62	6.17	4.70	5.10	4.55	4.90
		LOW	3.2	7.12	6.90	5.62	6.24	4.70	5.16	4.55	4.96
		HIGH	1.6	7.12	6.79	5.62	6.13	4.70	5.06	4.55	4.86
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	7.41	6.73	5.54	5.88	4.73	4.75	4.57	4.51
		LOW	3.6	7.41	6.79	5.54	5.93	4.73	4.80	4.57	4.56
		HIGH	2.0	7.41	6.69	5.54	5.83	4.73	4.70	4.57	4.47
	M 75%	AVERAGE	2.2	7.41	6.70	5.54	5.84	4.73	4.71	4.57	4.48
		LOW	3.2	7.41	6.77	5.54	5.91	4.73	4.77	4.57	4.54
		HIGH	1.6	7.41	6.66	5.54	5.80	4.73	4.68	4.57	4.44

TABLE C-28. LAND VOLUME REQUIRED FOR SPENT SOLIDS/ASH DISPOSAL, ACRE-FT/YR

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	0.35	1.73	0.87	4.33	0.81	8.67	0.62	11.56
		LOW	4.2	0.35	2.01	0.87	5.02	0.81	10.03	0.62	13.38
		HIGH	2.3	0.35	1.43	0.87	3.58	0.81	7.15	0.62	9.54
	I 85%	AVERAGE	2.9	0.35	1.59	0.87	3.99	0.81	7.97	0.62	10.63
		LOW	3.8	0.35	1.87	0.87	4.67	0.81	9.34	0.62	12.45
		HIGH	2.1	0.35	1.35	0.87	3.38	0.81	6.76	0.62	9.01
	M 78.7%	AVERAGE	2.5	0.35	1.45	0.87	3.63	0.81	7.26	0.62	9.68
		LOW	3.4	0.35	1.72	0.87	4.31	0.81	8.62	0.62	11.50
		HIGH	1.8	0.35	1.24	0.87	3.10	0.81	6.20	0.62	8.26
	SIP 56%	AVERAGE	1.0	0.35	0.92	0.87	2.29	0.81	4.59	0.62	6.11
		LOW	1.2	0.35	0.98	0.87	2.44	0.81	4.89	0.62	6.52
		HIGH	0.8	0.35	0.86	0.87	2.14	0.81	4.28	0.62	5.71
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	0.19	0.48	0.48	1.20	0.45	2.41	0.34	3.21
		LOW	3.7	0.19	0.54	0.48	1.35	0.45	2.70	0.34	3.60
		HIGH	2.0	0.19	0.43	0.48	1.07	0.45	2.14	0.34	2.86
	M 75%	AVERAGE	2.2	0.19	0.43	0.48	1.09	0.45	2.17	0.34	2.90
		LOW	3.2	0.19	0.50	0.48	1.25	0.45	2.50	0.34	3.34
		HIGH	1.6	0.19	0.40	0.48	0.99	0.45	1.98	0.34	2.63
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	0.22	0.49	0.55	1.23	0.51	2.47	0.39	3.29
		LOW	3.6	0.22	0.55	0.55	1.37	0.51	2.75	0.39	3.67
		HIGH	2.0	0.22	0.45	0.55	1.12	0.51	2.24	0.39	2.99
	M 75%	AVERAGE	2.2	0.22	0.46	0.55	1.14	0.51	2.28	0.39	3.04
		LOW	3.2	0.22	0.52	0.55	1.30	0.51	2.59	0.39	3.46
		HIGH	1.6	0.22	0.42	0.55	1.04	0.51	2.09	0.39	2.78

TABLE C-29. LAND VOLUME REQUIRED FOR SPENT SOLIDS/ASH DISPOSAL, HECTARE -m/yr

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW							
				8.8		22		44		58.6	
				CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC	CONVENTIONAL	AFBC
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	0.04	0.21	0.11	0.53	0.10	1.07	0.08	1.43
		LOW	4.2	0.04	0.25	0.11	0.62	0.10	1.24	0.08	1.65
		HIGH	2.3	0.04	0.18	0.11	0.44	0.10	0.88	0.08	1.18
	I 85%	AVERAGE	2.9	0.04	0.20	0.11	0.49	0.10	0.98	0.08	1.31
		LOW	3.8	0.04	0.23	0.11	0.58	0.10	1.15	0.08	1.54
		HIGH	2.1	0.04	0.17	0.11	0.42	0.10	0.83	0.08	1.11
	M 78.7%	AVERAGE	2.5	0.04	0.18	0.11	0.45	0.10	0.90	0.08	1.19
		LOW	3.4	0.04	0.21	0.11	0.53	0.10	1.06	0.08	1.42
		HIGH	1.8	0.04	0.15	0.11	0.38	0.10	0.76	0.08	1.02
	SIP 56%	AVERAGE	1.0	0.04	0.11	0.11	0.28	0.10	0.57	0.08	0.75
		LOW	1.2	0.04	0.12	0.11	0.30	0.10	0.60	0.08	0.80
		HIGH	0.8	0.04	0.11	0.11	0.26	0.10	0.53	0.08	0.70
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	0.02	0.06	0.06	0.15	0.06	0.30	0.04	0.40
		LOW	3.7	0.02	0.07	0.06	0.17	0.06	0.33	0.04	0.44
		HIGH	2.0	0.02	0.05	0.06	0.13	0.06	0.26	0.04	0.35
	M 75%	AVERAGE	2.2	0.02	0.05	0.06	0.13	0.06	0.27	0.04	0.36
		LOW	3.2	0.02	0.06	0.06	0.15	0.06	0.31	0.04	0.41
		HIGH	1.6	0.02	0.05	0.06	0.12	0.06	0.24	0.04	0.32
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	0.03	0.06	0.07	0.15	0.06	0.30	0.05	0.41
		LOW	3.6	0.03	0.07	0.07	0.17	0.06	0.34	0.05	0.45
		HIGH	2.0	0.03	0.06	0.07	0.14	0.06	0.28	0.05	0.37
	M 75%	AVERAGE	2.2	0.03	0.06	0.07	0.14	0.06	0.28	0.05	0.37
		LOW	3.2	0.03	0.06	0.07	0.16	0.06	0.32	0.05	0.43
		HIGH	1.6	0.03	0.05	0.07	0.13	0.06	0.26	0.05	0.34

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TABLE C-30. DOLLARS/kg SO_2 REMOVED

COAL TYPE	SULFUR CONTROL LEVEL AND PERCENTAGE REDUCTION	SORBENT REACTIVITY	CA/S RATIO	BOILER CAPACITY-MW			
				8.8	22	44	58.6
EASTERN HIGH SULFUR (3.5% S)	S 90%	AVERAGE	3.3	2.04	5.16	4.84	4.13
		LOW	4.2	2.87	6.13	5.72	[illegible]
		HIGH	2.3	1.11	4.27	3.86	3.15
	I 85%	AVERAGE	2.9	1.63	4.78	4.41	[illegible]
		LOW	3.8	2.47	5.72	5.29	4.58
		HIGH	2.1	0.84	4.06	3.65	[illegible]
	78.7%	AVERAGE	2.5	1.22	4.37	3.97	3.27
		LOW	3.4	2.05	5.17	4.85	4.14
		HIGH	1.8	0.56	3.74	3.29	2.58
	SIP 56%	AVERAGE	1.0	-0.35	2.85	2.25	[illegible]
		LOW	1.2	-0.17	3.03	2.43	1.79
		HIGH	0.8	-0.54	2.66	2.07	1.42
EASTERN LOW SULFUR (0.9% S)	S/I 83.9%	AVERAGE	2.8	-0.45	2.75	2.07	1.19
		LOW	3.7	-0.27	2.93	2.25	1.57
		HIGH	2.0	-0.62	2.59	1.91	1.05
	M 75%	AVERAGE	2.2	-0.59	2.61	1.93	1.06
		LOW	3.2	-0.39	2.82	2.13	1.26
		HIGH	1.6	-0.72	2.49	1.81	0.94
SUBBITUMINOUS LOW SULFUR (0.6% S)	S/I 83.2%	AVERAGE	2.7	-2.06	1.69	0.40	-0.51
		LOW	3.6	-1.88	1.87	0.58	-0.18
		HIGH	2.0	-2.20	1.56	0.27	-0.45
	M 75%	AVERAGE	2.2	-2.17	1.58	0.29	-0.42
		LOW	3.2	-1.47	1.78	0.49	-0.23
		HIGH	1.6	-2.29	1.47	0.18	-0.55

APPENDIX D

WESTINGHOUSE ESTIMATES OF AFBC INDUSTRIAL BOILER COST

Independent estimates of industrial AFBC boiler cost prepared by Westinghouse Research and Development Center are included in this Appendix. Values presented in terms of $\$/10^6$ Btu output were estimated by GCA based on total costs and boiler efficiencies presented by Westinghouse.

TABLE D-1. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (30×10^6 Btu/hr) - 150 psig SATURATED STEAM (SO_2 CONTROL LEVEL - STRINGENT)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	2272349.	2296855.	2313603.	2193840.	2205447.	2215856.	2192939.	2204377.	2214643.
WORKING CAPITAL	151391.	159442.	166326.	148659.	150058.	151558.	121131.	122471.	123908.
TOTAL CAPITAL COSTS	2423740.	2456297.	2479929.	2342499.	2355505.	2367414.	2314069.	2326849.	2338552.
FIXED ANNUAL COSTS, $/10⁶Btu		2.62			2.46			2.53	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	605563.	637767.	665305.	594638.	600233.	606231.	484522.	489885.	495633.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1080385.	1116972.	1147643.	1057724.	1065154.	1072822.	944724.	951891.	959282.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	207.	219.	230.	183.	185.	188.	185.	187.	190.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24509.	24365.	24001.	24903.	24880.	24805.	24273.	24251.	24179.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.39	8.72	9.10	8.08	8.15	8.23	7.40	7.47	7.55
$/10⁶Btu output	8.16	8.49	8.85	7.86	7.93	8.01	7.21	7.27	7.35

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-2. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (30×10^6 Btu/hr) - 150 psig SATURATED STEAM (SO_2 CONTROL LEVEL - INTERMEDIATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%		3.50			.50			.60	
COAL HHV, BTU/LB		11800.			13800.			9600.	
% REMOVAL OF SO2 RELEASED		85.00			84.67			84.00	
SO2 EMISSIONS, LBS/MM BTU		.89			.20			.20	
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	2265162.	2286558.	2304558.	2193840.	2205447.	2215856.	2192939.	2204377.	2214643.
WORKING CAPITAL	145367.	155837.	162673.	148659.	150058.	151558.	121131.	122471.	123908.
TOTAL CAPITAL COSTS	2414469.	2442395.	2467231.	2342499.	2355505.	2367414.	2314069.	2326849.	2338552.
FIXED ANNUAL COSTS, $/10^6 Btu		2.60			2.46				2.52
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	597226.	623349.	650691.	594638.	600233.	606231.	484522.	489885.	495633.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1070790.	1100690.	1131344.	1057724.	1065154.	1072822.	944724.	951891.	959282.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	203.	213.	224.	183.	185.	188.	185.	187.	190.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24554.	24444.	24099.	24903.	24880.	24805.	24273.	24251.	24179.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.30	8.57	8.93	8.08	8.15	8.23	7.40	7.47	7.55
$/10^6 Btu output	8.07	8.34	8.69	7.86	7.93	8.01	7.21	7.27	7.35

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSER QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-3. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (30×10^6 Btu/hr) - 150 psig SATURATED STEAM (SO_2 CONTROL LEVEL - MODERATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	2255199.	2276367.	2295290.	2186604.	2198391.	2209380.	2185801.	2197423.	2208264.
WORKING CAPITAL	146655.	152560.	159169.	147946.	149175.	150575.	120447.	121624.	122966.
TOTAL CAPITAL COSTS	2401854.	2428927.	2454458.	2334549.	2347565.	2359955.	2306248.	2319048.	2331230.
FIXED ANNUAL COSTS, $/10^6 Btu		2.59			2.45			2.52	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	586619.	610240.	636675.	591784.	596699.	602299.	481787.	486498.	491865.
OVER HEAD	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.	127920.
TOTAL ANNUALIZED COST	1058464.	1085766.	1115624.	1053742.	1060502.	1067846.	940875.	947404.	954468.
PERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KW	199.	208.	219.	182.	184.	186.	184.	186.	188.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	24612.	24506.	24184.	24917.	24895.	24827.	24287.	24265.	24201.
STEAM COST, $/1000 LB (@ 60% LOAD)	8.18	8.43	8.78	8.05	8.10	8.18	7.37	7.43	7.50
$/10^6 Btu output	7.96	8.20	8.54	7.83	7.89	7.96	7.17	7.23	7.30

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BLSSER QUARRY
-III (LOW REACTIVITY) - MEALO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-4. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE BY AFBC INDUSTRIAL BOILER (75×10^6 Btu/hr) - 150 psig SATURATED STEAM (SO_2 CONTROL LEVEL - STRINGENT)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%		3.50			.60			.60	
COAL HHV, BTU/LB		11800.			13800.			9600.	
% REMOVAL OF SO2 RELEASED		90.00			84.67			84.00	
SO2 EMISSIONS, LBS/MM BTU		.59			.20			.20	
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	3978153.	4016202.	4044345.	3857191.	3873529.	3887879.	3856458.	3872517.	3886638.
WORKING CAPITAL	296764.	316892.	334103.	289936.	293433.	297182.	221114.	224465.	228058.
TOTAL CAPITAL COSTS	4274917.	4335094.	4378448.	4147127.	4166962.	4185061.	4077572.	4096982.	4114697.
FIXED ANNUAL COSTS, $/10^6 Btu		1.83			1.73			1.77	
OPERATING COSTS & 60% LOAD									
TOTAL DIRECT OPERATING COST	1187056.	1267567.	1336411.	1159744.	1173732.	1188728.	884456.	897861.	912232.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	1986873.	2075243.	2145626.	1941217.	1957941.	1975406.	1658940.	1675025.	1691817.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	516.	546.	574.	458.	464.	470.	464.	469.	474.
STEAM GENERATED & 100% OPERATING LOAD, LBS/HR	61273.	60912.	60003.	62258.	62199.	62011.	60683.	60627.	60447.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.17	6.48	6.82	5.93	5.99	6.06	5.20	5.26	5.33
$/10^6 Btu output	6.00	6.31	6.63	5.77	5.83	5.90	5.06	5.12	5.18

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - PEALC QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-5. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE BY AFBC INDUSTRIAL BOILER (75×10^6 Btu/hr) - 150 psig SATURATED STEAM (SO_2 CONTROL LEVEL - INTERMEDIATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	3966910.	4001425.	4029297.	3857191.	3873529.	3887879.	3856458.	3872517.	3886638.
WORKING CAPITAL	291554.	307881.	324969.	289936.	293433.	297182.	221114.	224465.	228058.
TOTAL CAPITAL COSTS	4258464.	4309306.	4354266.	4147127.	4166962.	4185061.	4077572.	4096982.	4114697.
FIXED ANNUAL COSTS, $/10^6 Btu		1.82			1.73			1.77	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1166215.	1231523.	1299877.	1159744.	1173732.	1180728.	884456.	897861.	912232.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	1963870.	2035849.	2109981.	1941217.	1957941.	1975406.	1658940.	1675025.	1691817.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	508.	533.	560.	458.	464.	470.	464.	469.	474.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61385.	61110.	60248.	62258.	62199.	62011.	60683.	60627.	60447.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.09	6.34	6.66	5.93	5.99	6.06	5.20	5.26	5.33
$/10^6 Btu output	5.92	6.17	6.48	5.77	5.83	5.90	5.06	5.12	5.18

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN SC% CAL
-II (MEDIUM REACTIVITY) - BLSSER QUARRY
-III (LOW REACTIVITY) - PEALO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-6. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE BY AFBC INDUSTRIAL BOILER (75×10^6 Btu/hr) - 150 psig SATURATED STEAM (SO_2 CONTROL LEVEL - MODERATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	3951566.	3985013.	4014019.	3847244.	3863601.	3878645.	3846668.	3862759.	3877569.
WORKING CAPITAL	284925.	299687.	316209.	288152.	291224.	294724.	219405.	222349.	225703.
TOTAL CAPITAL COSTS	4236490.	4284701.	4330229.	4135397.	4154826.	4173369.	4066073.	4085107.	4103272.
FIXED ANNUAL COSTS, $/10^6 Btu		1.81			1.72			1.77	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1139698.	1198750.	1264838.	1152609.	1164897.	1178898.	877619.	889395.	902812.
OVER HEAD	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.	189330.
TOTAL ANNUALIZED COST	1934449.	1999861.	2071835.	1932452.	1947436.	1963982.	1650903.	1664922.	1680837.
PERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KW	498.	520.	547.	456.	460.	466.	461.	466.	471.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	61529.	61265.	60461.	62293.	62237.	62068.	60717.	60663.	60501.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.98	6.21	6.52	5.90	5.95	6.02	5.17	5.22	5.29
$/10^6 Btu output	5.82	6.04	6.35	5.74	5.79	5.86	5.03	5.08	5.14

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-7. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (150×10^6 Btu) - 450 psig, 600 F STEAM (SO_2 CONTROL LEVEL - STRINGENT)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	7488914.	7551400.	7590633.	7309500.	7331652.	7350662.	7309296.	7331006.	7349668.
WORKING CAPITAL	499903.	540158.	574581.	486247.	493241.	500739.	348603.	355306.	362491.
TOTAL CAPITAL COSTS	7988817.	8091558.	8165214.	7795747.	7824893.	7851401.	7657895.	7686312.	7712159.
FIXED ANNUAL COSTS, $/10^6 Btu		1.72			1.64			1.68	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1995613.	2160633.	2298323.	1944987.	1972964.	2002956.	1394412.	1421223.	1449965.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3429274.	3603444.	3750303.	3347089.	3378999.	3412516.	2782719.	2813370.	2845556.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	1033.	1093.	1148.	917.	927.	939.	927.	937.	949.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111086.	110431.	108784.	112872.	112765.	112425.	110017.	109914.	109588.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.87	6.21	6.56	5.64	5.70	5.78	4.81	4.87	4.94
$/10^6 Btu output	5.18	5.48	5.79	4.98	5.03	5.09	4.25	4.30	4.36

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN SC% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - PEALC QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-8. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (150×10^6 Btu) - 450 psig, 600 F STEAM (SO_2 CONTROL LEVEL - INTERMEDIATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%		3.50			.50			.60	
COAL HHV, BTU/LB		11800.			13800.			9600.	
% REMOVAL OF SO2 RELEASED		85.00			84.67			84.00	
SO2 EMISSIONS, LBS/MM BTU		.89			.20			.20	
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	7472069.	7525269.	7566816.	7309500.	7331652.	7350662.	7309296.	7331006.	7349668.
WORKING CAPITAL	489483.	522137.	556314.	486247.	493241.	500739.	348603.	355306.	362491.
TOTAL CAPITAL COSTS	7961491.	8047425.	8123129.	7795747.	7824893.	7851401.	7657895.	7686312.	7712159.
Fixed Annual costs, $/10⁶ Btu		1.71			1.64			1.63	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1957931.	2086546.	2225254.	1944987.	1972964.	2002956.	1394412.	1421223.	1449965.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3384082.	3525742.	3671931.	3347089.	3378999.	3412516.	2782719.	2813370.	2845556.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	1017.	1065.	1120.	917.	927.	939.	927.	937.	949.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111290.	110791.	109229.	112872.	112765.	112425.	110017.	109914.	109588.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.79	6.05	6.40	5.64	5.70	5.76	4.81	4.87	4.94
$/10⁶ Btu output	5.10	5.34	5.64	4.98	5.03	5.09	4.25	4.30	4.36

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN SC% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - MENLO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-9. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (150×10^6 Btu) - 450 psig, 600 F STEAM (SO_2 CONTROL LEVEL - MODERATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	7445258.	7499990.	7542812.	7296367.	7318198.	7337968.	7296400.	7317820.	7337239.
WORKING CAPITAL	476224.	505750.	538794.	482680.	488824.	495824.	345184.	351072.	357781.
TOTAL CAPITAL COSTS	7925482.	8005740.	8081605.	7779047.	7807022.	7833792.	7641585.	7668893.	7695020.
Fixed Annual Costs, $/10^6 Btu		1.70			1.63			1.68	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	1904896.	2023000.	2155175.	1930719.	1955295.	1983295.	1380737.	1404290.	1431123.
OVER HEAD	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.	286290.
TOTAL ANNUALIZED COST	3326400.	3454864.	3596595.	3330546.	3358924.	3390511.	2766820.	2794089.	2824428.
PERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KW	997.	1041.	1094.	911.	920.	932.	922.	931.	942.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	111550.	111072.	109614.	112935.	112834.	112528.	110077.	109980.	109687.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.67	5.92	6.24	5.61	5.66	5.73	4.78	4.83	4.90
$/10^6 Btu output	5.00	5.22	5.51	4.95	5.00	5.06	4.22	4.26	4.32

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 98% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - PEALO QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-10. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (200×10^6 Btu) - 750 psig, 750 F STEAM (SO_2 CONTROL LEVEL - STRINGENT)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	90.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.59			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.83	3.41	5.26	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	10023556.	10100137.	10147459.	9808234.	9833745.	9855388.	9808444.	9833412.	9854632.
WORKING CAPITAL	654254.	707928.	753824.	636046.	645371.	655369.	452521.	461458.	471038.
TOTAL CAPITAL COSTS	10677810.	10808065.	10901283.	10444280.	10479117.	10510757.	10260964.	10294870.	10325671.
FIXED ANNUAL COSTS, $/10⁶Btu		1.73			1.65			1.69	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2617017.	2831711.	3015297.	2544183.	2581486.	2621475.	1810082.	1845830.	1884153.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4526241.	4757484.	4552568.	4420150.	4462109.	4506258.	3667727.	3708014.	3750393.
PERFORMANCE									
BOILER EFFICIENCY, %	83.94	83.45	82.21	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	1377.	1457.	1531.	1222.	1236.	1252.	1237.	1250.	1265.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	139562.	138739.	136669.	141805.	141671.	141244.	138218.	138089.	137680.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.17	6.52	6.89	5.93	5.99	6.07	5.05	5.11	5.18
$/10⁶Btu output	5.13	5.42	5.73	4.93	4.98	5.05	4.20	4.25	4.31

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - PEALC QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-11. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (200 × 10^6 Btu) - 750 psig, 750 F STEAM (SO_2 CONTROL LEVEL - INTERMEDIATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	85.00			84.67			84.00		
SO2 EMISSIONS, LBS/MM BTU	.89			.20			.20		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.50	2.94	4.68	2.42	2.85	4.57	2.42	2.85	4.57
CAPITAL COSTS									
TOTAL TURNKEY	10003149.	10068167.	10118102.	9806234.	9833745.	9855388.	9808444.	9833412.	9854632.
WORKING CAPITAL	640360.	683899.	729468.	636046.	645371.	655369.	452521.	461458.	471038.
TOTAL CAPITAL COSTS	10643509.	10752066.	10847570.	10444280.	10479117.	10510757.	10260964.	10294870.	10325671.
FIXED ANNUAL COSTS, $/10^6 Btu		1.72			1.65			1.69	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2561441.	2735595.	2517873.	2544183.	2581486.	2621475.	1810082.	1845830.	1884153.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4466297.	4654297.	4848422.	4420150.	4462109.	4506258.	3667727.	3708014.	3750393.
PERFORMANCE									
BOILER EFFICIENCY, %	84.10	83.72	82.54	85.29	85.21	84.96	83.14	83.06	82.81
AUXILIARY POWER, KW	1356.	1421.	1494.	1222.	1236.	1252.	1237.	1250.	1265.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	139817.	139191.	137228.	141805.	141671.	141244.	138218.	138089.	137680.
STEAM COST, $/1000 LB (@ 60% LOAD)	6.08	6.36	6.72	5.93	5.99	6.07	5.05	5.11	5.18
$/10^6 Btu output	5.05	5.29	5.59	4.93	4.98	5.05	4.20	4.25	4.31

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - PEALC QUARRY
SORBENT PARTICLE SIZE - 500. MICRONS

TABLE D-12. ESTIMATED CAPITAL, OPERATING COSTS AND PERFORMANCE OF AFBC INDUSTRIAL BOILER (200×10^6 Btu) - 750 psig, 750 F STEAM (SO_2 CONTROL LEVEL - MODERATE)

COAL TYPE	EASTERN HIGH S			EASTERN LOW S			WESTERN SUBBITUMINOUS		
COAL SULFUR CONTENT, WT%	3.50			.90			.60		
COAL HHV, BTU/LB	11800.			13800.			9600.		
% REMOVAL OF SO2 RELEASED	79.77			75.00			75.00		
SO2 EMISSIONS, LBS/MM BTU	1.20			.33			.31		
SORBENT TYPE	I	II	III	I	II	III	I	II	III
CA/S MOLAR RATIO	2.09	2.51	4.13	1.92	2.33	3.87	1.92	2.33	3.87
CAPITAL COSTS									
TOTAL TURNKEY	9975848.	10037312.	10086601.	9793307.	9818255.	9840670.	9793805.	9818251.	9840242.
WORKING CAPITAL	622682.	662050.	706108.	631290.	639482.	648815.	447562.	455813.	464758.
TOTAL CAPITAL COSTS	10598530.	10699362.	10794710.	10424597.	10457737.	10489485.	10241767.	10274064.	10305000.
FIXED ANNUAL COSTS, $/10^6 Btu		1.71			1.64			1.69	
OPERATING COSTS @ 60% LOAD									
TOTAL DIRECT OPERATING COST	2490729.	2646200.	2624834.	2525158.	2557927.	2595261.	1791850.	1823253.	1859031.
OVER HEAD	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.	380360.
TOTAL ANNUALIZED COST	4389830.	4560213.	4748340.	4398470.	4435700.	4477240.	3646902.	3682659.	3722542.
PERFORMANCE									
BOILER EFFICIENCY, %	84.30	83.93	82.83	85.34	85.27	85.03	83.18	83.11	82.89
AUXILIARY POWER, KW	1329.	1388.	1458.	1215.	1227.	1242.	1230.	1241.	1256.
STEAM GENERATED @ 100% OPERATING LOAD, LBS/HR	140145.	139544.	137712.	141885.	141757.	141373.	138294.	138172.	137804.
STEAM COST, $/1000 LB (@ 60% LOAD)	5.96	6.22	6.56	5.90	5.95	6.03	5.02	5.07	5.14
$/10^6 Btu output	4.95	5.17	5.45	4.90	4.95	5.01	4.17	4.22	4.27

COSTS IN MID 1978 DOLLARS
SORBENT TYPE -I (HIGH REACTIVITY)- WESTERN 90% CAL
-II (MEDIUM REACTIVITY) - BUSSEN QUARRY
-III (LOW REACTIVITY) - PEALC CLARRY
SORBENT PARTICLE SIZE - 500. MICRONS